Biomass Gasification: Fast Internal Circulating Fluidised Bed Gasifier Characterisation and Comparison

A thesis submitted in fulfilment of the requirements for the Degree of Master of Engineering in Chemical and Process Engineering

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Abstract

In 2004 the Department of Chemical and Process Engineering (CAPE) at University of Canterbury began a programme to investigate using biomass gasification integrated combined cycle (BIGCC) technology to convert waste products and residues to useful energy for the wood processing sector. This research was conducted as a part of Objective Two of the programme to develop gasification and gas cleaning technology. This project involved commissioning and characterising the operation of the Fast Internal Circulating Fluidised Bed (FICFB) gasifier and comparing its operation with a more conventional up-draught process owned and operated by Page Macrae in Mount Manganui. The wood derived gas composition of each gasifier was measured using gas chromatography and these compositions were used to calculate lower heating values (LHV).

The CAPE FICFB gasifier has proven to produce successfully a gas with a lower heating value of 10400-12500 kJ/Nm^3. The Page Macrae gasification process produces a low quality gas with a lower heating value of 4100-5100 kJ/Nm^3. This is much lower than the CAPE gasifier since the oxidant used in the up-draught gasification process is air and the product gas is diluted by nitrogen. The Page Macrae gasification system combusts wood derived gas to produce steam for a laminar veneer lumber (LVL) processing plant so gas quality and heating value are less important than in electrical production applications. Reducing the nitrogen content of the CAPE product gas will increase the heating value of the gas. Improvements to the boiler system will reduce the amount of air required for gasification and hence reduce the nitrogen content. Further improvements to gas quality can be gained from a change in the fuel feed point from on top of the gasification column's bubbling fluidised bed to the side of the bubbling fluidised bed. The CAPE gasifier is much more complicated and requires specialised operators but produces a gas suitable for gas engine and gas turbine technology. Overall the CAPE gasification system is more suited to BIGCC applications than the Page Macrae process

Acknowledgements

There are many people who have had an influence on the out come of not only my masters but also the CAPE gasification project and they all deserve many thanks but the following have had a major influence for me in particular.

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1 Introduction

1.1 Renewable Electricity Supply for New Zealand

Status of New Zealand Electricity Supply

New Zealand's Electricity supply has been predominantly based on large scale hydro and natural gas for the past three decades. The make up of the fuel type for electricity supplied in 2004 is shown below in Table 1-1 (Dang H 2005). Although renewable fuels make up the majority of the supply outlined below, the projected growth of demand of 2% per annum in the period to the end of 2012 will stretch the resources available. The cancellation of project Aqua, a 524 MW hydro project, in 2004 has seen an increase in interest in new coal, wind, geothermal and gas projects to make up the projected short fall (Concept Consulting Group 2004).

Table 1-1 Electricity generation by fuel type 2004

Fuel Type	Energy [PJ]	% of Total Generation
Hydro	94.5	63.9
Gas[1]	23.8	16.1
Geothermal[1]	9.5	6.4
Coal[1]	14.4	9.7
Oil	0.1	0.0
Wind	1.7	1.1
Others[1,2]	4.0	2.7
TOTAL	**147.9**	**100**

[1] Includes output of cogeneration plants for these fuel types.
[2] Includes electricity generation from biogas, waste heat and wood sources.

Kyoto Protocol

The Intergovernmental Panel on Climate Change (IPCC) has predicted an average global rise in temperature of 1.4°C to 5.8°C between 1990 and 2100 (Faure, Gupta et al. 2003). The effects this rise in temperature will have on the climate are only speculative but include more extreme weather patterns and a rise in sea level rises. The Kyoto Protocol was adopted in 1997 under the United Nations Framework Convention on Climate Change in an attempt to

reduce the effects of global warming. The Protocol legally binds signatories to reduce the emissions of six greenhouse gases (including carbon dioxide and methane) collectively to 5 % below 1990 levels by 2008-2012. The ratification of the Protocol by Russia in 2004 satisfied the clause which required 55 % of world carbon dioxide emissions in 1990 to be accounted for the Protocol to come into force (Faure, Gupta et al. 2003). Between 1990 and 2004 New Zealand's total carbon dioxide equivalent green house gas (GHG) emissions increased by 17.8% as shown in a sector break down in Table 1-2. The emissions from electricity production increased 73 % over the period or 4 % per annum as seen in Table 1-3 (M.E.D. 2005). In Table 1-3 the reduction in emissions from natural gas is due to a reduction in supply as the Maui gas field comes off line. However, this is outweighed by the increase in coal usage to make up for this decline and to meet increasing demand. Initial estimates predicted that the growth in carbon dioxide emissions in New Zealand would be absorbed by increased growth in plantation forestry. However, as Table 1-2, shows this is not the case, although there is a noticeable increase in the absorption from land use and forestry.

Table 1-2. Carbon dioxide equivalent emissions for various sectors (Petrie, Wear et al. 2006). The energy sector includes electricty, transport and heating. Agriculture includes emissions from animals and derived wastes.

Sector	kt CO_2 equivalent		Change from 1990 (kt CO_2 equivalent)	Change from 1990 (%)	Average change per year over period
	1990	2004			
Energy	23,655	31,647	7,992	33.8	2.4
Industrial processes	3,214	4,202	987	30.7	2.2
Solvent and other product	41	48	6	16.4	1.2
Agriculture	32,116	36,866	4,750	14.8	1.1
Land-use change and forestry	-18,977	-24,482	-5504	29.0	2.1
Waste	2,482	1,839	-642	-25.9	-1.9
Total	**42,532**	**50,122**	**7,590**	**17.8**	**1.3**

Table 1-3. CO2 Emissions from electricity generation in kt of CO2 equivalent.

Year	Coal	Liquid	Gas	Total
1990	474	14.7	2997	3486
2004	3746	1.6	2297	6041

The structure of the Protocol allows for renewable energy projects that displace existing GHG emissions to gain tradable carbon credits and a financial windfall for reducing emissions. If as a country we decide to keep our commitments to the Kyoto protocol, the predicted investment in new coal and other non-renewable sources of energy could have long term detrimental effects for the economy as well as the environment. Biomass fuel comes under the renewable definition and therefore has an economic as well as ecological benefit and may provide a solution to the problem.

Biomass as a fuel

In global terms biomass ranks fourth as an energy resource making up 14 % of the world's primary energy needs. In developing nations, biomass is an even more important resource providing as much as 35 % of the energy needs in some areas of the globe, particularly in isolated areas where it is often the only resource available (Hall, Rosill-Calle et al. 1992). Much of the current use of biomass in the developed world is heavily subsidised to make it possible for the resource to compete with much cheaper fossil fuels. Woody biomass wastes and residues are often used on commercial wood processing and paper manufacturing sites to generate heat and electricity for on site use; however, if biomass is going to make a major contribution to the world's energy mix then dedicated energy crops are the only answer. By 2010 large quantities (~5 EJ) of high yielding energy crops could compete economically with coal in some areas of the USA (Bain, Overend et al. 1998). There are no large scale trials being undertaken to substantiate these claims but these will be required to obtain the necessary investment.

In 2004 the Ministry of Agriculture and Forestry reported the New Zealand to have 1,822,000 ha of managed commercial forest from which 20 million m3 of round-wood is harvested annually (Li and Pang 2005). Harvesting this round-wood results in around 4 million m3/yr of waste residues and processing results in a further 4.5 million m3/yr in waste residues (Li and Pang 2005). The thermal energy content of the residues amounts to approximately 33 GW, more than New Zealand's total installed electricity generation capacity, leaving much scope for carbon neutral energy generation.

1.2 Canterbury University Gasification Program

In 2004 the Biomass Energy Group in the Department of Chemical and Process Engineering began a four-year research program to investigate biomass integrated gasification combined cycle (BIGCC) technology for New Zealand. The ultimate goal of the program is to produce electricity generating systems for the wood processing industry using waste wood by-products. This involves adapting existing BIGCC technology to utilise the waste biomass supply currently available in New Zealand. The project consists of four separate objectives:

1. Evaluation of BIGCC technologies developed overseas. The aim is to recommend a BIGCC system for the process design and modelling and identify areas where further technology development is needed. The systems reviewed include the biomass handling, gasifier type, gasification agent, gas cleaning and gas turbine

2. Transfer and development of BIGCC system to suit NZ conditions. This builds on the outcomes of the technology review taking the recommendations and putting them into the design of a laboratory scale gasifier, to be used to undertake research and development. This encompasses biomass handling and feed systems, and gasifier reactor and gas cleaning technology. This thesis fits under objective two.

3. Modelling of feedstock supply and energy demand. The modelling of the energy demand of large wood processing plants and modelling of the biomass supply. The first part involves energy modelling of wood processing plants to assess the suitability and scale of the technology required to meet their demands. The second part involves assessing the biomass available, where it is, what form it is in and the cost of transporting it to a site where it can be utilised.

4. Design and modelling of BIGCC systems. This objective involves developing a model of the gasifier using operating parameters to predict performance and assess the economics for the different gasifier configurations, fuel types and scales.

1.3 Gasification

Producing gas from carbonaceous material is by no means a new idea: the first commercial coal-fuelled plant began operation in 1812 in England, producing coal gas for lighting and cooking as a by-product of coke production for the metallurgical industry via a pyrolysis type process. This initial process paved the way for gasification producing gas as a source of light and heat. Higman and van der Burgh (2003) define gasification as the "production of gases with a usable heating value from carbonaceous fuels" (p1). Since the 1920s advances in technology have seen producer gas consisting of carbon monoxide and hydrogen used in the production of chemicals and liquid fuels. Recent advances in gasifier, gas turbine and combined cycle technology mean that gasification also has more economic applications as a source of electrical power. Outside New Zealand large petroleum coke and coal integrated gasification combined cycle plants are already in commercial operation. The Nuon Power Buggenum Integrated Gasification Combine Cycle (IGCC) power plant in the Netherlands has been operating since 1983. In 2002 the plant began operating using biomass as a secondary fuel, since then it has operated at up to 30% biomass and plans are to operate the plant at up to 50% biomass (Power Retrieved March 2007). Biomass gasification processes offers electrical efficiencies of 35-40% as opposed to efficiencies of 15-28% for the conventional biomass steam turbine cycles (Franco and Giannini 2005).

Thermodynamics of Gasification

From the most basic point of view gasification is successful because a gas is much easier to handle than solid fuel; moreover there are thermodynamic reasons illustrated by the following reaction sets:

$C + 1/2\,O_2 = CO$	-111 MJ/kmol
$CO + 1/2\,O_2 = CO_2$	-283 MJ/kmol
$C + O_2 = CO_2$	-394 MJ/kmol

If approximately 30 % of the energy available in the carbon is used forming carbon dioxide in the exothermic two step reaction above, the remaining carbon can be converted into combustible gaseous product. The gaseous compounds formed are methane, carbon monoxide and heavier hydrocarbons. Taking the hydrogen in the fuel into account increases the conversion even further with modern processes converting 75% to 90% of their heating value

into gas (Higman and van der Burgt 2003). The gasification process can be broken down into a series of steps: drying, devolitisation, char gasification and gas phase reactions.

Drying

The first stage of the processes is the driving off of moisture bound to the biomass.

Devolitisation

These are low temperature endothermic reactions that occur between 350°C and 800°C and involve the volatile component of the biomass fuel. The reactions take place simultaneously with heating and the rate and products formed are dependant on the temperature. At low temperatures, below 500°C, the reactions are mainly gasification but as the temperatures tend toward 800°C and above they become more and more pyrolytic. The products resulting from these reactions include: char (fixed carbon), tars (described later), light gases such as methane, carbon monoxide, carbon dioxide, hydrogen and steam (Smoot and Smith 1985). The products of the initial reactions then crack and oxidise as they interact with other gaseous molecules.

Char Gasification

The char gasification reactions occur in parallel with the drying and devolitisation steps. Char gasification is a range of rate limiting reactions for carbon conversion. The reactions are several orders of magnitude slower than the drying and devolitisation steps (Bridgwater 1995). Dry woody biomass pellets have a carbon content of 51 wt % (dry basis) so the char reactions are very important. The char reactions are catalysed by the presence of potassium , sodium and calcium bound to the char (Clemens, Gong et al. 2004). Three reactions govern the gasification of char they are:

(1) Boudouard Reaction

$C + CO_2 = 2CO$ +172 MJ/kmol

(2) Hetrogeneous Water gas Shift Reaction (Steam gasification)

$C + H_2O = CO + H_2$ +131 MJ/kmol

(3) Methanation

$C + 2H_2 = CH_4$ -75 MJ/kmol

The Boudouard and water gas shift reactions are several orders of magnitude faster than the methanation reactions. Char gasification reactions are very slow in biomass gasification because of the large amounts of carbon monoxide and hydrogen gas formed in the initial devolitisation reactions and also because of the relatively low temperatures. The temperatures of biomass gasification are limited by the high moisture content of the fuels and also by the relatively low melting point of the ash, when compared with coal. The ash melting point is low due to the high concentrations of potassium compounds (21 wt% of ash) (Appendix A provides detailed results of ash analysis).

Homogeneous Gas Phase Reactions

Homogeneous gas phase reactions are used to describe the changes in gas composition as the gases interact after the initial devolitisation and char gasification steps. Tars formed in the devolitisation are also thermally cracked and hydro-cracked in this phase of reactions. Reactions (4) and (5) describe the interactions of carbon monoxide and methane with steam in the gas phase:

(4) Homogeneous Water gas Shift Reaction

$CO + H_2O = CO_2 + H_2$ -41 MJ/kmol

(5) Steam Methane Reforming Reaction

$CH_4 + H_2O = CO + 3H_2$ +206 MJ/kmol

Biomass Derived Synthesis Gas to Electrical Power

Almost all installed commercial methods of electricity generation from biomass are either stand alone steam turbine or co-generation facilities, but more advanced and efficient processes do exist. The USA has over 1000 wood-fired plants with over 7500 MW of installed electric capacity because of Federal subsidies introduced in the 1970s (Williams and Larson 1996). Only a third of these actually sell electricity commercially: most are owned by wood products and paper manufacturing industries and are small scale, typically 10-25 MW (Bain, Overend et al. 1998). Because the energy density of biomass is much less than coal and gas and requires much more effort to collect, economic plants are limited in size to about 100 MW. Most of these plants are very primitive, having efficiencies in the range of 14-25 % HHV and do not even use modern boiler design or materials in order to reduce capital

investment for the relatively small size to make them more economic. Ultra-super critical steam turbine (Rankin cycle) technology can achieve efficiencies of the order of 46 % owing to the peak steam temperatures being limited to the order of 600°C at 30 MPa based on coal with a lower heating value (LHV) of 25 MJ/kg (Rosenkranz and Wichtmann 2006). There has been little improvement in this over the past four decades although there is scope to increase peak steam cycle temperature. Recent advances in super alloys have made 350 bar and 700°C possible and pulverized fuel demonstration plants are planned to show efficiencies greater than 50% (LHV) are possible operating with coal (Rosenkranz and Wichtmann 2006).

Gas turbine technology uses the hot exhaust gases in various ways to attain very high cycle efficiencies. The peak gas temperature of a gas turbine cycle is in excess of 1200°C and research is leading to continued increases in the peak temperature and the overall efficiency of the cycles. Since World War Two large investment in jet engine technology has led to an annual average increase of cycle temperatures of 20°C. These advances are being incorporated into emerging technology in the form of the aero-derivative turbines with cycle efficiencies approaching 50 % (LHV) (Beer, Grant et al. 2002). Combined cycle systems on the market today achieve efficiencies in excess of 50 % operating on natural gas, which is the most efficient available. The efficiency of IGCC plants which incorporate combine cycle technology operate with an overall plant efficiency of around 41.6% (LHV) which is similar to super critical steam combustion plants operating with coal at 41.3% (LHV) (Beer, Grant et al. 2002). The problem with applying IGCC technology in New Zealand is that it is very complicated and very sensitive to fuel quality, requiring specialised maintenance and expertise to operate.

Other applications for the use of the product gas include gas engines, operating on the Otto (spark ignition) cycle or a modified Diesel (injection ignition) cycle with dual fuels. Although engines are less sensitive than gas turbines to tars and particulates, tars condensed on valves and harden when a plant is shut down and cooled, restricting the valves from fully closing when restarted. Engines have lower efficiencies than turbines but are also cheaper and smaller in scale, and are thus more likely to be used in small scale distributed generation. Maintenance of these engines is also less specialised. Gasification also lends itself to heat and boiler applications, where it provides similar advantages to staged combustion. The heat applications mean that combustor size is reduced resulting in reduced capital investment. Low excess air leads to higher efficiency, reduced nitrogen oxides and less fly ash fouling in the

boiler or heat exchanger. In New Zealand, the gasification systems have the potential to replace natural gas as the fuel in existing gas fired boiler systems as the source of fuel with little or no modification (Coulter 2005).

1.4 Gasifier Reactor Types

The reactor systems being used through-out the world to gasify biomass are dominated by four configurations which are counter current moving bed, co-current moving bed, entrained bed and fluidised bed. In this thesis only the moving bed and fluidised bed configurations that operate at atmospheric pressure are reviewed as only these two types are relevant to this work. High pressure reactors are operated around the world; however, the complexities related to the increased pressure means they require specialised operation and maintenance, beyond the scope of what is available on most wood processing sites in New Zealand. The moving or fixed bed reactor consists of a bed of biomass sitting on top of a moving grate and the oxidant is fed through it either from above or below. The fluidised bed configuration consists of biomass being fed into a bed of sand fluidised with oxidant at high temperatures.

Counter Current Moving Bed (Up-draught)

The counter current moving bed processes are the oldest type of gasification process. Biomass fuel is fed through the top of the reactor on to a packed bed which gravitates down as the fuel below is gasified. The oxidant is introduced through the bottom of the reactor, moving up through the packed bed of fuel as shown in Figure 1-1. Oxidants include oxygen, air or a mixture of air and steam. Wet biomass is initially dried at the top of the bed in the drying zone as hot product gases pass through it. The dried biomass gravitates down as the fuel below is used up and new fuel is fed on top. The dry biomass devolatilizes in the distillation or pyrolysis zone to form light gases, tars and solid char at temperatures between 350°C and 800°C. The solid char is partially gasified in the reduction zone to form light gases. Most of the remaining char is combusted in the hearth zone which sits on top of a moving grate. Any remaining ash and char material falls through the grate and is carried away by the removal system. The temperature of the hearth zone dictates the extent of combustion of the char; however, this temperature is restricted by the softening point of the ash. The ash softening point of biomass can be as low as 800°C. The types of grate used are many and varied, serving the purpose of keeping the fuel porous for oxidant flow and fuel reactivity.

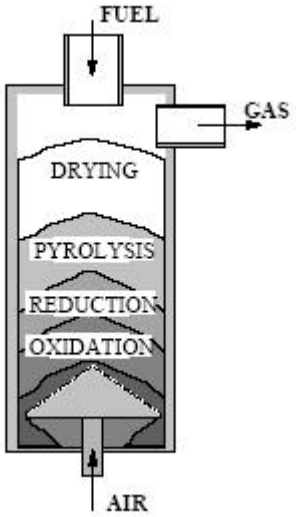

Figure 1-1 Counter-current moving bed gasification configuration. Oxidant moves up through packed bed of fuel gravitating down.

This type of gasification process is very simple and well understood and has many advantages because of the vast experience many companies have operating these processes on large commercial scales. Commercial scale electricity plants were constructed in the South America and Europe during the 1940s and 1950s based on the counter current moving bed configuration. Most of these plants have been shut down because of pollution created by waste water used to remove ash and char from the bottom of the bed and to remove tar and particulate material from the raw gas. The low gas exit temperature resulting from the product gases passing through the wet biomass, thereby acting as an internal heat exchanger, combined with high char burnout leads to high gas conversion efficiency. As a result of the internal heat exchanger mechanism, the biomass fuel can be of high moisture content, up to 60 % (wet basis). The same mechanism also means that volatile gases are not subject to high temperatures and so little cracking of tar material occurs and a dirty gas results. This is not a problem for heat and steam applications where the product gases are combusted directly in a boiler. The dirty gases generated require much more cleaning for applications in engines or turbines. Up-draught gasification systems typically produce a gas with a calorific value of 4-6 MJ/Nm^3 with air as the oxidant or 8-14 MJ/Nm^3 with pure oxygen (Bridgwater 1995). The tar content ranges from 50 g/Nm^3 to 150 g/Nm^3 (Milne, Evans et al. 1998).

Co-Current Moving Bed (Down-draught)

In the co-current moving bed configuration, a packed bed of fuel moves down under gravity as new fuel is fed from above and as fuel below is gasified (see Figure 1-2). Oxidant is fed either from the top or mid-way down the reactor. The zones in the bed are much the same as the counter current configuration with new fuel drying on the top of the bed then moving down into the pyrolysis or distillation zone, where the biomass devolatilizes. The drying and pyrolysis zones are heated by a mixture of radiation and convection from the hearth zone below. Pyrolysis gases are partially combusted in the hearth zone together with some of the char formed in the pyrolysis reactions and the gases reach temperatures of approximately 1200°C depending on the exact design. Below the hearth zone is the reduction zone where char sits on top of the grate and this is where the char gasification reactions occur.

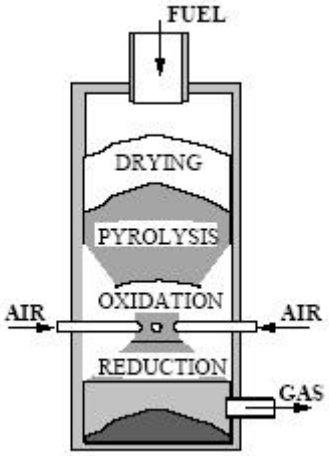

Figure 1-2, Co-current moving bed gasification configuration. Fuel is fed to the top of the reactor, oxidant is fed to the middle of the bed and the product gases leave the reactor at the bottom.

The down-draught gasification system is characterised by a hot low tar product gas because of the gas passes through the hot oxidation zone. Although the product gases often have low tar content they are never tar free, probably because some of the gases formed do not passing through the hottest regions and have a low residence time in the oxidation zone. The gases produced also contain high levels of because the product gases pass through the reduction zone before leaving the reactor. The high product gas temperature leads to a lower overall efficiency than the up-draught gasification configuration. Fuel moisture content is limited because the drying and pyrolysis zones are heated by radiation and convection from below rather than forced convection in the case of the counter current process. Fuel moisture levels

are limited to 25 % (wet basis) since drying occurs because of natural convection and radiation. The throat, which separates the pyrolysis zone from the hearth zone, limits the fuel size and means that only uniform sized fuel can be used, to prevent blockages. The throat also limits the scale up possibilities of this system. The calorific value of the gas produced in down-draght gasification systems is very dependant on the oxidant used: the calorific value is 4 to 6 MJ/Nm³ with air or 9-11 MJ/Nm³ with oxygen (Bridgwater 1995). The tar content of down-draght gasification systems ranges from 0.1- 2.4 g/Nm³.

Entrained Bed (Slagging) Gasifier

The entrained bed gasifier is widely used in coal gasification, but has had limited applications with woody biomass to date because a liquid or fine particulate fuel is required for the process. The Shell gasification used in chemical and IGCC applications is based on this type of process. Fuel and oxidant are fed into the top of the reactor chamber at high temperatures, above 1300°C. The high temperature melts the ash to form a molten slag which runs down the walls and is captured in the bottom. The high temperatures results in high carbon conversion and low tar concentration. Oxygen is normally used as the oxidant so a relatively high calorific value gas can be generated, but this is also a limiting factor as large scale operations are required to counter the additional capital cost of an oxygen production plant. The process only operates on liquid and small particulate fuels so fuel can require pre-treatment. All solid fuels require grinding to small diameters. A German process uses fast pyrolysis of straw biomass to produce oil containing char particles (oil-char slurry) to feed an entrained bed gasifier used in chemical and liquid fuels production plants (Dinjus June 2006).

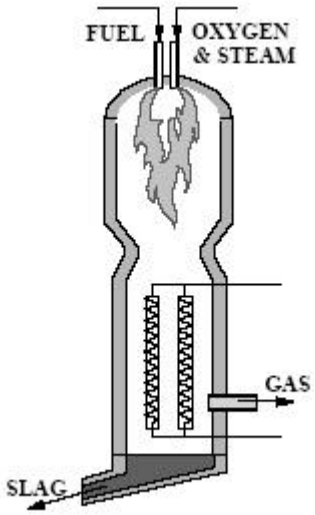

Figure 1-3. Entrain bed gasifier, fuel and oxidant enter reactor co-currently.

Fluidised Bed Gasification

Fluidised bed reactors offer many features not available from other types of reactor including the high heat and mass transfer rates and good mixing of the solid phase. These features lead to high overall reaction rates and uniform bed temperatures, hence consistent product gas. The fluidised bed reactor consists of a bed of sand material with particulate sizes in the order of 0.1 to 5 mm fluidised with air, oxygen or steam as oxidant. Biomass fuel is fed into the bed which operates at temperatures between 700 and 900°C, lower than the peak temperatures in fixed bed gasification. The bed temperature in fluidised bed gasifiers is limited by the ash softening point temperature above which the ash is sticky. Above this point agglomeration of the bed material can occur, causing lumps which block the fluidisation.

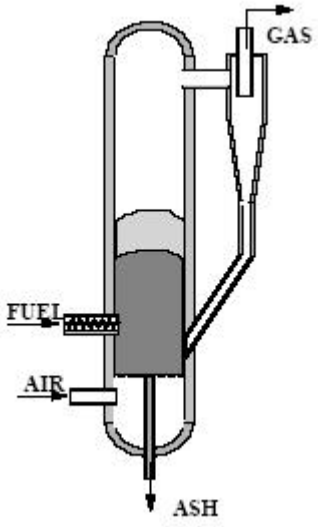

Figure 1-4. Fluidised bed gasifier. The type of fluidised bed is determined by the superficial fluid velocity, high velocity generates a circulating bed, low velocity a bubbling bed.

The superficial velocity of the fluidising gas determines the characteristics of the bed. Bubbling fluidised beds have low superficial gas velocity. Some of the fine bed material and ash is entrained in the gas and separated out in a cyclone. One characteristic of the bubbling fluidised bed system is low char conversion because of the low residence time of the fuel and slow reactivity of the char. Circulating fluidised beds have a higher gas velocity which entrains some of the bed material and char in the gas flow. The entrained material is separated from the gas stream and circulated back to the main reactor. The high velocities of the recirculation system and violent mixing lead to a higher overall carbon conversion. The circulating bed reactor is currently used for wood waste conversion in the pulp and paper industry and for firing cement and lime kilns. The first commercial scale BIGCC plant in

Varnamo, Sweden was a high pressure circulating fluidised bed design. Fluidised bed gasification systems produce gases with low to moderate calorific values similar to up-draught gasification systems, 4-6 MJ/Nm³ with air, 8-14 MJ/Nm³ with oxygen. Fluidised bed configuration also allows steam to be used as an oxidant, producing a gas with a calorific value of between 12 MJ/Nm³ and 20 MJ/Nm³ (Bridgwater 1995).

Fast Internal Circulating Fluidised Bed (FICFB) Reactor

The FICFB reactor is also often referred to as a dual fluid bed reactor; however, this is a less specific terminology. The basic idea of this reactor is to separate the gasification and combustion and increase the calorific value of the product gas without using oxygen (Hofbauer, Veronik et al. 1997). Biomass fuel is fed into the gasification zone where it devolatilizes and gasifies in the presence of steam fed through the bottom of the bubbling fluidised bed. Residual char from these reactions circulates with bed material into the combustion zone where it is combusted with air in a circulating fluidised bed, heating the bed material. Combustion gases are separated from the hot bed material, which is circulated into the gasification zone, without mixing of the combustion and gasification product gases, providing the heat for the endothermic gasification reactions.

The use of steam as the oxidant increases the calorific value of the gas produced which avoids the dilution effects of nitrogen in air and without the capital cost of an oxygen plant. However the gasification process in effect is essentially a pyrolytic one and hence the tar concentration is high (Bridgwater 1995). The char circulation mechanism leads to high overall carbon conversion and high efficiency. Like all fluid bed systems, this design is subject to agglomeration of bed material caused by ash softening. The separation of combustion and gasification systems also leads to a high hydrogen concentration in the product gas meaning that the process can also be used for toward Fischer Tropsch liquid fuels production and hydrogen separation in the longer term. Dolomite lime or limestone can be used in the bed material to absorb carbon dioxide, further increasing the hydrogen concentration and reducing the need for downstream processing (Hofbauer, Veronik et al. 1997).

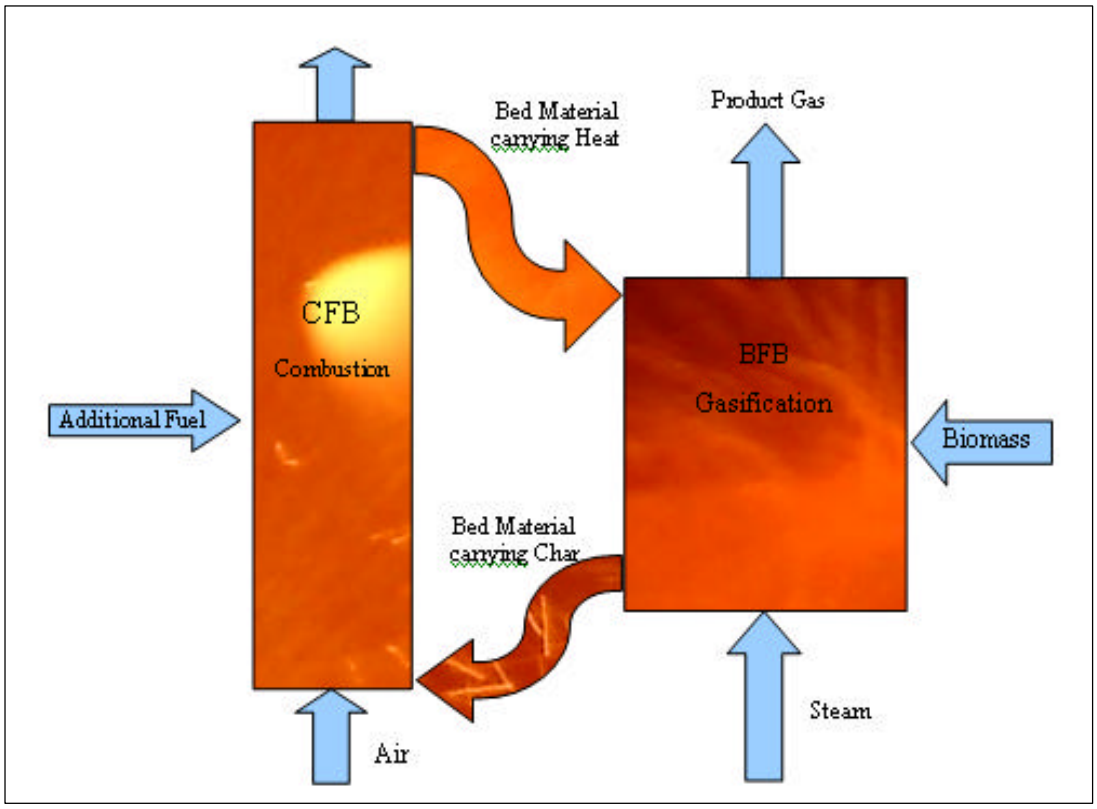

Figure 1-5. The Fast Internal Circulating Fluidised Bed (FICFB) gasification system.

The FICFB process has been proven on a large scale at Gussing in Austria. A FICFB gasifier was built to use waste wood from the wood industry in the surrounding area and produces 2 MW_{el} of electricity in a gas engine and heat for a community heating programme. The fuel input rating of the gasifier is 8 MW_{th}. The gas produced is high in hydrogen and carbon monoxide and has a calorific value of the order of 13 MJ/Nm^3 (Hofbauer, Reinhard et al. 2002). The tar concentration of the product gases operating under optimised conditions is 5 g/Nm^3 (Hofbauer, Veronik et al. 1997).

Technology Comparison

A review of technology undertaken in objective 1 by (Li and Pang 2005) assessed the BIGCC processes operated around the world and the success of the individual projects. Six major projects were identified as having BIGCC technology that was applicable to the New Zealand situation. Table 1-4 shows a comparison of the gas composition produced by typical gasification reactor configurations for gas composition and gas quality. The major projects assessed were:

- ARBRE Project, Yorkshire, UK
- Bioelettrica, Italy
- Greve in Chianti, Italy
- Varnamo, Sweden
- RRB in Andhro Pradesh, India
- Gussing, Austria

Table 1-4. Comparison of different gasification reactors types adapted from (Bridgwater 1995) and (Hoffbauer, Reinhard et al. 2002)

Reactor Type	Gas Composition (vol %, dry)					HHV (MJ/m^3)	Gas Qualuity	
	H_2	CO	CO_2	CH_4	N_2		Tars	Particulates
Fluid bed air Blown	9	14	20	7	50	5.4	Fair	Poor
Updraught, air-blown	11	24	9	3	53	5.5	Poor	Good
Downdraugh, air-blown	17	21	13	1	48	5.7	Good	Fair
Downdraught, oxygen-blown	32	48	15	23	0	10.4	Good	Good
Multi-solid fluid bed	15	47	15	23	0	16.1	Fair	Poor
Twin fluidised bed gasification (FICFB).	40	25	20	10	5	13.0	Fair	Poor
Pyrolysis	40	20	18	21	1	13.3	Poor	Good

The FICFB gasification process used at the Gussing plant in Austria was identified as having the most potential for developing a successful BIGCC project. The Austrian developed FICFB plant at Gussing has been successfully operated since 2001 with over 10000 hours in 2004, much longer than any of the other projects reviewed in this study. Fixed bed and entrained bed systems have been extensively developed elsewhere so further work on these configurations was considered counter productive. The multi-solid fluidised bed has not been developed beyond the bench scale and development to commercial scale would require work outside the time frame of the CAPE program. The researchers identified product gas clean up and feed handling as two areas where attention is required for further investigation (Li and Pang 2005).

1.5 Tars and Contaminants

The IEA Biomass Gasification Task 33 meeting in Brussels in 1998 stated that: "All organics boiling at temperatures above that of benzene should be considered as Tar". Tars formed in the gasification of woody biomass are a complex mixture of condensable hydrocarbons derived from the reaction of the lignin. The compounds formed are mainly aromatic rings ranging in size from single rings to five member rings; they also contain oxygenated compounds such as phenols and poly-nuclear aromatic hydrocarbons (PAHs). The compounds formed are dependent on a number of factors including the type of gasifier reactor configuration and reaction kinetics involved. From an operational point of view tars are any compounds that condense on surfaces of the gasifier or downstream process equipment; however, this is not enough for modern processes. Milne (1998) states "the organics produced under thermal or partial oxidation regimes (gasification) of any organic material are called tars and are generally assumed to be largely aromatic".

Tars are well known to be the "Achilles heel" of biomass gasification but there are also other unwanted contaminants that require removal before the product gas is clean enough for further use. Table 1-5 shows the gas requirements for gas turbines and gas engines. The chlorine and sulphur content of biomass is low so they are not a problem. Product gases need to be cooled to remove particulates as high temperature removal systems such as cyclones and ceramic filters are not developed to the stage where they are as efficient as low temperature methods (Milne, Evans et al. 1998).

Table 1-4. Gas quality requirements for commercially available gas engine and turbine technology (Scharf and Carrington 2005), converted to Nm^3 basis.

	Guascor Engines	Jenbacher Engines	Gas Turbines
Particles >5microns	0 mg/Nm^3	Not Specified	10ppm
Particles <5microns	108 mg/Nm^3	Not Specified	
Tars	108 mg/Nm^3	5 mg/Nm^3	5mg to 5g/Nm^3
Sulphur	2520 mg/Nm^3	700 mg/Nm^3	1 to 7000 ppm
Ammonia	54 mg/Nm^3	50 mg/Nm^3	50 ppm
Chlorine	126 mg/Nm^3	100 mg/Nm^3	0.5 ppm
Silicon	7.2 mg/Nm^3	200 ppm	Not Specified

Gas cleaning or tar removal and reduction systems are categorised into two types, primary where contaminants are reduced in the gasifier itself and secondary where contaminants are removed from the product gases. Primary methods of tar reduction either prevent tar formation or convert the tars in the reactor itself. These methods involve designing the reactor properly and optimising the operating parameters of the reactor, using additives or catalysts to help favourable reactions take place. Secondary methods can involve using separate reactors to crack tar material thermally or catalytically. Mechanical methods such a cyclones, baffle filters, ceramic filters, fabric filters, electrostatic filters and scrubbers are also classed as secondary methods.

Primary reduction of tar formation in the gasification reactor itself tends to involve changing operating conditions or using bed additives to prevent catalytically their formation and crack any tars formed. One of the parameters often used is the bed temperature of the reactor, for fluid bed gasification temperatures above 800°C are optimal and higher free board temperatures also have a favourable impact on tar concentration (Devi, Ptasinski et al. 2003). Many studies show that temperature increases have a positive impact on tar concentrations: a temperature increase of 700°C to 900°C reduced tar concentrations by 40 % in a free fall pyrolysis reactor (Yu, Brage et al. 1997). Raising the bed temperature from 700°C to 850°C reduced tars by 74 % and also reduced ammonia concentrations in a bubbling fluidised bed gasification process (Narvaez, Orio et al. 1996). The tar concentration of a product gas from a FICFB dual fluidised bed steam gasifier being operated in Vienna, Austria was reduced from 7 g/Nm³ to 2 g/Nm³ with a temperature increase of 800°C to 880°C. Increased heat losses and reduced plant efficiency are draw backs associated with increased reactor temperature. The reactor temperature in the moving and fluidised bed configurations is limite by the ash fusion temperature. Many in bed catalysts have been trialled for a range of different purposes including tar reduction. These additives have a range of effects including influencing gas composition and heating value, promoting char gasification, reducing tar formation and can also prevent solids agglomeration. Different catalysts that are reported to have a favourable effect include nickel-based catalysts of many types, calcined dolomites, magnesites, zeolites, olivine and iron based catalysts (Devi, Ptasinski et al. 2003). A nickel/olivine catalyst was developed at the University of Strasbourg in France. When it was used in a 100 kW FICFB gasifier, in Vienna, the tar concentration of the product gas was reduced by up to 43% (Pfeifer, Rauch et al. 2004).

The oxidant to fuel ratio also affects the tar concentration of the gas produced, but has the undesirable effect of reducing the heating value of the gas as more of the gas is combusted. When a FICFB gasifier similar to the one being developed at the University of Canterbury was used in Vienna, an increase in the biomass fuel to steam ratio from 0.2 to 0.5 (steam/fuel by weight) resulted in the tars being reduced from about 7 to 4 g/Nm^3 (Hofbauer and Rauch 2000). Fuel composition and moisture content also affect the tar concentration and composition of the product gas. High moisture content fuels reduce bed temperatures and increase the tar yield.

1.6 Project Aims

Construction and Commissioning

The Chemical and Process Engineering (CAPE) FICFB gasifier was designed by Rick Dobbs in 2004 and construction began the same year. Initially it was hoped this would be completed by the time this research began in January 2005, but for a variety of reasons the work took longer than expected. The initial thesis proposal put together in 2004 did not allow for this delay, and hence the final structure of this differs from the proposal. The construction of the columns and commissioning of the plant is described in the following sections of this thesis. Commissioning included the cold testing of the fluidisation characteristics of the columns and bed material, and also included developing the LPG burner system used to heat the beds to operating conditions. The aim of commissioning the plant was to develop an operating procedure to be followed for any subsequent operation (The operating procedure is described in the gasifier operations guide attached as Appendix C).

Characterisation of Process Parameters

Characterising the product gas of the CAPE gasifier first included design and set up of the gas cleaning and analysis system which is described in the experimental section. The aim of this research was to operate the gasifier under a range of conditions and assess the effects of each on the product gas composition and other parameters. The results from this experimentation were then used to compare the operation with the up-draught gasification system owned and operated by Page Macrae in Mount Manganui. The Page Macrae process was characterised by changing process parameters and monitoring their affects on other parts of the plants operation.

2 Construction

The FICFB gasifier at Canterbury University was designed under contract by Rick Dobbs, a chemical and process engineer who works for Lemar Environmental. The circulating fluidised bed, bubbling fluidised bed and stand pipe and top bed material transfer section were designed based on the work of (Grace, Avidan et al. 1997). The cyclones were designed based on (Hoffmann and Stein 2002). Construction began in late 2004 with Rick Dobbs and the Chemical and Process Engineering Department technical staff. In January 2005 the author began work towards an ME and started working along side Rick Dobbs on construction. Rick Dobbs worked full time on the gasifier construction and commissioning until October 2005; however, he continued to have a major role working part time gradually reducing hours and handing over responsibility. When the plant HAZOP took place December 2005, the gasifier had been operated at high temperatures greater than 800°C but not for extended periods. In 2006 commissioning of the plant continued undertaken by the author with help from Jack Rutherford and Rick Dobbs acting as a consultant on a casual basis as required. Design and setup of the analysis systems was the full responsibility of the author.

2.1 Process Description

The construction of CAPE FICFB gasifier can be broken down in several different parts: the fuel feed system, gasification column, combustion column, the bed material transfer chute, top bed material transfer system and the afterburner. Figure 2.1 shows a process flow diagram of the design material and energy balance of the CAPE gasification system. The gasifier plant process and instrumentation diagram is in Appendix B.

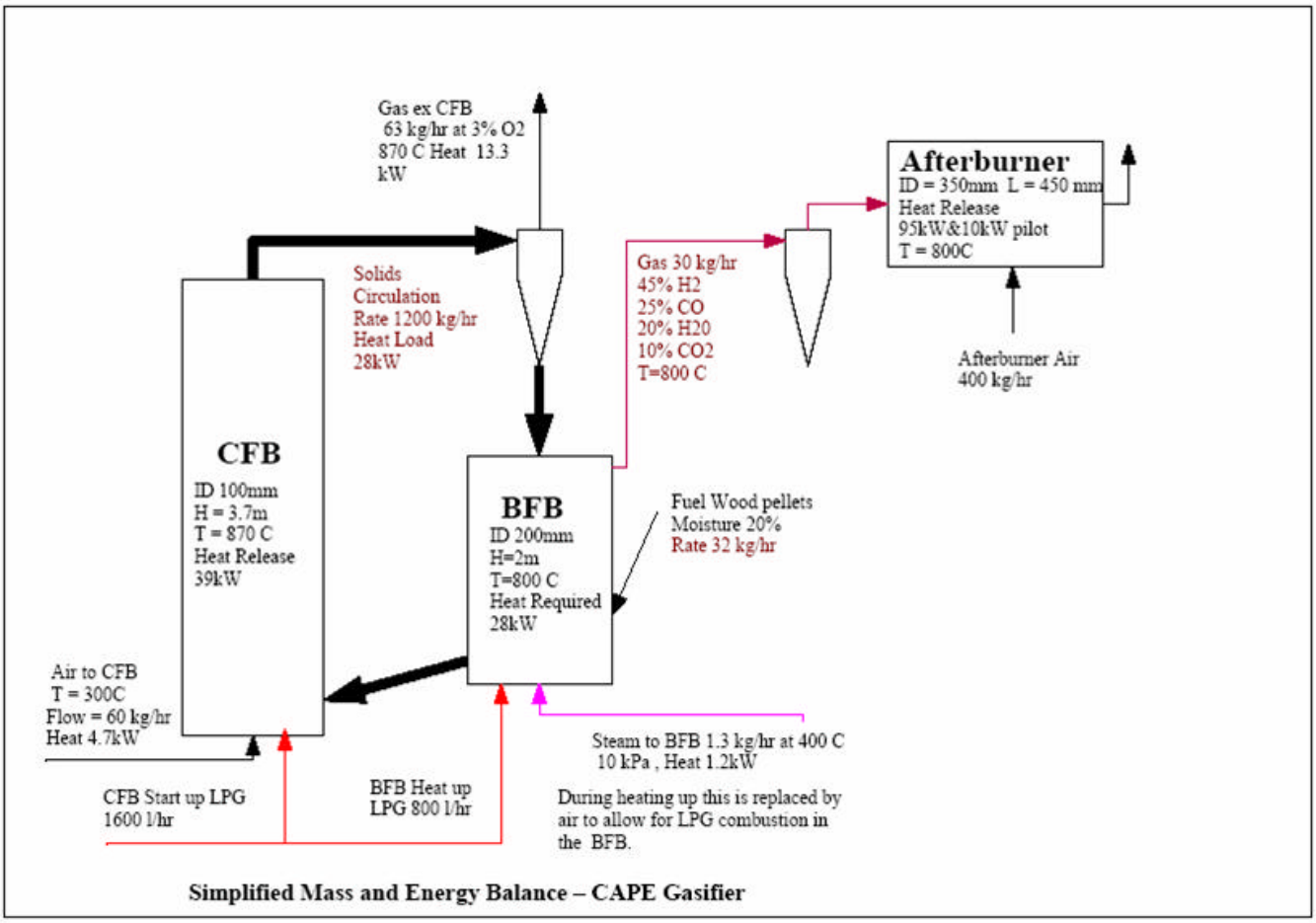

Figure 2-1. Process flow diagram and mass balance of Canterbury University FICFB gasifier

Fuel feed system

The FICFB gasification system incorporates two parallel fluidised beds to gasify fuel and combust residual char material. Biomass fuel is fed to a bubbling fluidised bed in a screw conveyor and drops on to the top of the bubbling bed. Woody fuel material in the form of pellets or chips (see Appendix A for details on composition) is fed to the auger via a sealed hopper. The fuel feed rate is controlled by a variable frequency (VF) drive connected to the

three phase electric motor running the screw. During operation the fuel feed rate is set and the other variables are manipulated to attain the desired set points.

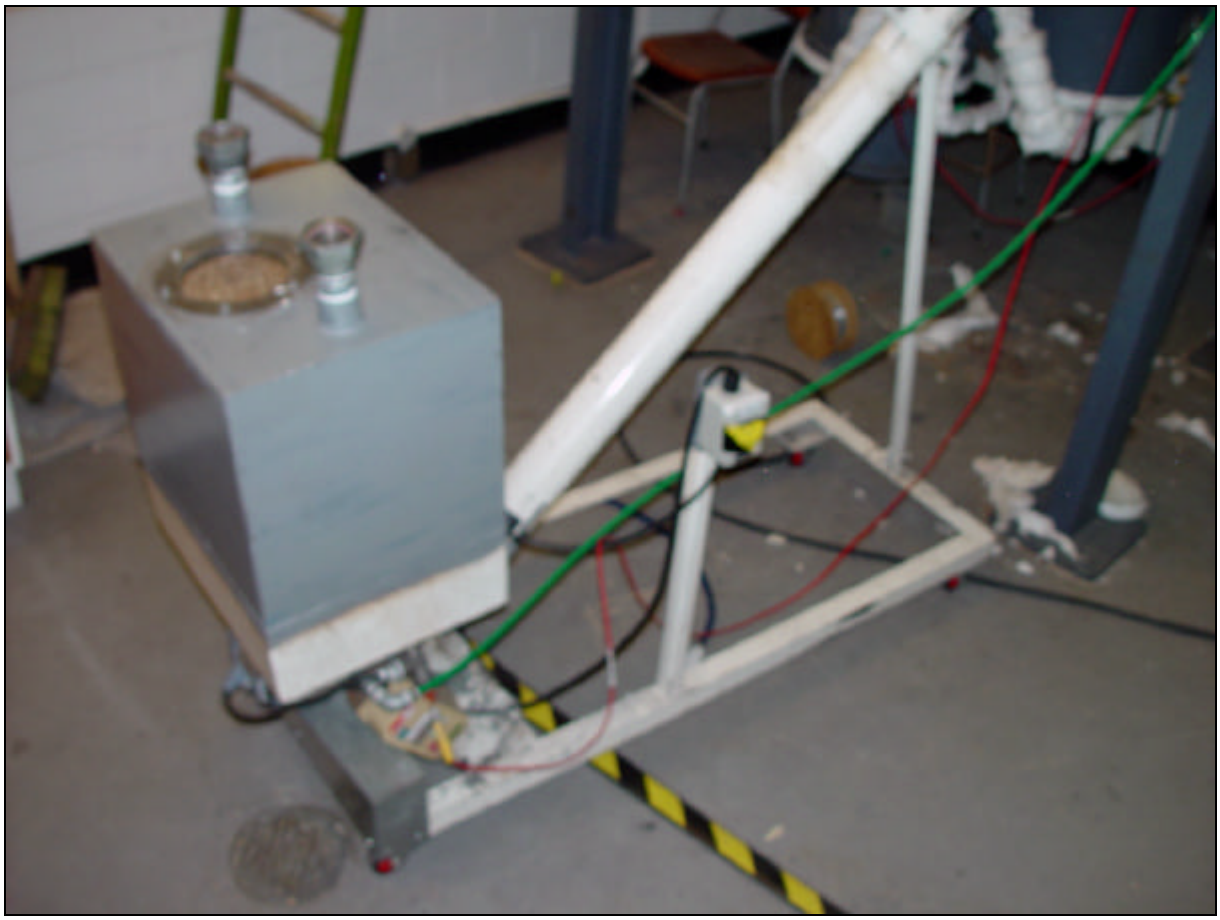

Figure 2-2. Feed Auger with the sealed hopper system.

The hopper is fitted with viewing ports so the fuel level can be inspected and holds approximately 60 kg of pellet material, enough for 2-3 hours' operation under gasification conditions. The hopper is refilled manually using a sealed rotary valve. To help prevent the back flow of combustible and toxic gases from the gasification column whilst the hopper is being refilled, a solids flow valve fitted at the top of the feed auger is closed. During operation nitrogen is slowly fed into the hopper so the hopper is constantly under positive pressure to prevent back flow of toxic and combustible gas from the gasification column. This keeps the auger temperature low, preventing the pellets from swelling and protects the auger screw and bearings from the effects of heat. The purge system would not be necessary on a completely sealed large scale plant. A typical nitrogen purge gas feed rate is 10 L/min.

The Gasification Column

Biomass fuel is fed on to the top of bed material in the gasification column where it reacts with steam in a bubbling fluidised bed. The reactions occurring here are described in the previous section and form the product gases and char. The bed material used is greywacke river sand screened to a diameter less than 1mm. Steam is fed to the bottom of the column to react with the wood and fluidise the bed material. The steam rate and hence oxidant to fuel ratio is controlled manually with a valve on the water feed rate to the boiler. During test operation the bed temperature is maintained between 700°C and 900°C but can be controlled with the wood feed rate, bed material circulation rate and circulating bed material temperature.

The column is constructed in a series of sections each with steel outer structure and lined with vermiculite insulation and refractory hot face. The sections are bolted together with 25 mm ceramic fibre (Kaowool) blanket gaskets made to fit over the insulation and refractory part of the column, allowing for expansion. The sections are sealed gas tight with high temperature room temperature vulcanising (RTV) sealant which forms gaskets (Ados gasket 260) on the steel flanges. The internal diameter of the column is 207 mm but the bottom section is cone shaped angled towards the chute where bed material and char circulate to the combustion column (see Figure 2-4), (detailed drawings are available in the gasifier folder found on the Engineering computer network).

Figure 2-3. The distributor plate for the gasification column with protruding nozzles. The nozzles are offset because of the cone shaped lower region of the column.

The column was designed so that bed material is fluidised with the product gas and steam; however, no allowances were made for the density difference between the char and bed material, this is discussed in further detail in Section 5 FICFB Gasifier Results and Discussion. Steam is fed into the bed in stainless steel distributor plates which have seven nozzles protruding from the main plate (Figure 2-3). Each nozzle consists of six 3 mm holes which are evenly spaced around the nozzle facing downward to reduce bed material flowing inside. The bed is heated to operating temperature with high pressure LPG fuel. The fuel is premixed with air in the distributor and injected into the bed through the nozzles in the distributor plate. The main fuel and air rates are controlled manually with valves to maintain a desired oxygen concentration. The fuel air mixture is ignited with a pilot flame situated above the bed surface. The pilot flame operates a mixture of low pressure LPG and compressed air which fires about half way across the column. The flow-rate and mixture of LPG and compressed air for the pilot burner is controlled with regulators on the air and LPG supplies. The flame is monitored with a flame management system using flame rods to detect a micro scale current which can be induced with a large electrical potential difference through the ionised propane and butane in the LPG during combustion.

Figure 2-4. Bottom section of the gasification column. The chute where bed material exits can be seen at the top of the photograph and the stainless steel distributor plate can be seen in the bottom.

Bed Material Transfer Chute

The mixture of bed material and char in the bubbling fluid bed of the gasification zone is transferred to the combustion zone via the bed material transfer chute or chute for short. The chute opens into the gasification zone just above the top of the nozzles in the gasification column (see Figure 2-4). The chute is cast of refractory in two halves, one half in the bottom section of the gasification column and the other in the bottom section of the combustion column, which can be seen in Figure 2-5. Bed material flows down the chute under gravity but this can be fluidised with either steam or air to promote bed material flow. Two fluidising gas distributor tubes (or sparge pipes) can be seen in Figure 2-5. These evenly distribute gas along the length of the chute to promote bed material flow. Each sparge pipe is 20 mm diameter with eight 1.5 mm holes evenly spaced along the length.

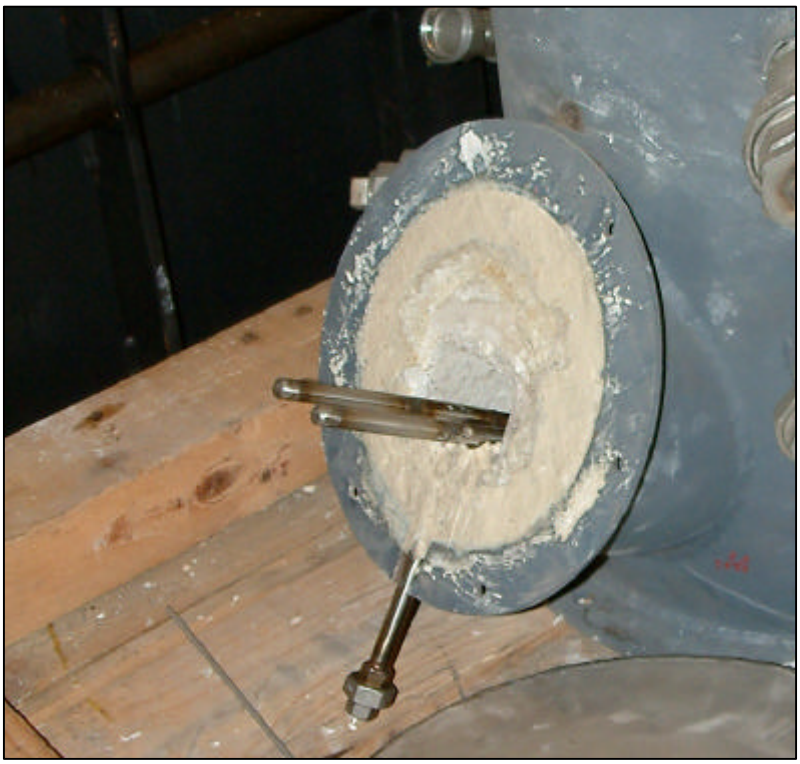

Figure 2-5. The bottom half of the chute which opens into the combustion column. The fluidising gas distributor tube can be seen on the bottom of the chute.

Bed material is heated in the combustion column by combusting char transferred in bed material from the gasification column and additional LPG fuel in a circulating fluidised bed. This is fluidised with air and combustion gases. The circulating bed requires high air rates because bed material is entrained and rises up the centre of the column then falls down the outside in a circulating motion. Fluidising air enters the column at three different heights. Fluidising air enters through nozzles in the distributor plate at the bottom of the bed at between 5 and 12 kg/h. Primary air is fed to the bottom section of the bed 250 mm above the distributor plate. It enters through four nozzles, each adjustable to allow the jet angle to moved up or down or parallel to the bed. The primary air rate is 2-5 kg/h. Secondary air is fed to the bed above the primary air 200 mm above the distributor plate. In the bottom section of the combustion column where bed material enters through the chute, gas velocities are limited to avoid leakage of gases into the chute and gasification column. The superficial gas velocities in this section are low and it behaves like a bubbling fluidised bed enabling bed material transfer from the chute into the riser column. The primary air supply allows the bed material to be transported vertically upwards without back flow of gas up the chute. The secondary air supply limits the formation of nitrogen oxide compounds and increases the upward velocity of the gases and bed material (Loffler, Kaiser et al. 2003). Loffler et al

(2003) describe the fluid dynamics and models the behaviour of the circulating bed used to combust char.

Figure 2-6. The bottom section of the combustion circulating fluidised bed column. The chute can be seen entering the column from the left hand side. The distributor plate can be seen in foreground and primary air inlets can be seen near bottom with the secondary inlet above them.

The column is constructed in two halves; the bottom half being sections of similar in construction to the gasification column and the top half being one stainless steel pipe section. The top half of the column is insulated with Kaowool blanket. The internal diameter of the column is 107 mm in both halves. The combustion column is heated to operating temperature using high pressure LPG fuel injected into the fluidised bed through the distributor plate. The bed has a pilot burner continuously burning in the circulating bed which ignites LPG injected through the distributor. The LPG flame controller operates by detecting the flame of the pilot burner with a flame rod located inside the burner tube and protected from the bed material.

Top Return System

The top return system transfers hot bed material from the combustion column to the gasification column to provide the heat for the endothermic gasification reactions. This section is one of the key features of the dual bed FICFB gasification system as it stops product gases flowing back out of the exhaust with the combustion product gases but allows the hot bed material to be transferred between the combustion column and the gasification column. The return system of the transfer duct, cyclone, stand pipe and siphon system can be seen in Figure 2-7. It has been designed to transfer up to 1200 kg/h of bed material at temperatures of 1000°C (See mass and energy balance calculations in Appendix D).

The first part in the bed material transfer system is the duct which exits the top of the combustion column and transfers bed material and hot combustion gases to the cyclone. The duct has a square cross section as it exits the combustion column but merges into a narrow rectangular shape as it enters the cyclone. The cyclone is a semi-scrolled design with a vortex tube fitted to the outlet to reduce the effects of particle bounce. It separates the hot combustion gases from the bed material; the bed material dropping into the stand pipe with the combustion gases exit to the exhaust system through heat exchangers that pre-heat the air.

Figure 2-7. The top solids transfer system. The combustion column can be seen to the right the the duct exiting at the top. The cyclone connects to the siphon at its bottom exit and the air preheating heat exchangers to the top left. The figure shows the system before it was insulated with Kaowool blanket.

The key to transferring bed material without back flow of product gases into the exhaust system lies in the stand pipe and siphon system. Bed material falls into the stand pipe from the cyclone forming an almost solid bed in the pipe, creating a large plug, preventing gases from flowing up the pipe. It then gravitates across the bottom section of the siphon with the assistance of small amounts of fluidising steam or air. The fluidising air and steam are concentrated under the upward section creating a fluidised bed in this section which allows the bed material to drop over into the return section leading to the gasification column as more material feeds to the fluid bed from the stand pipe.

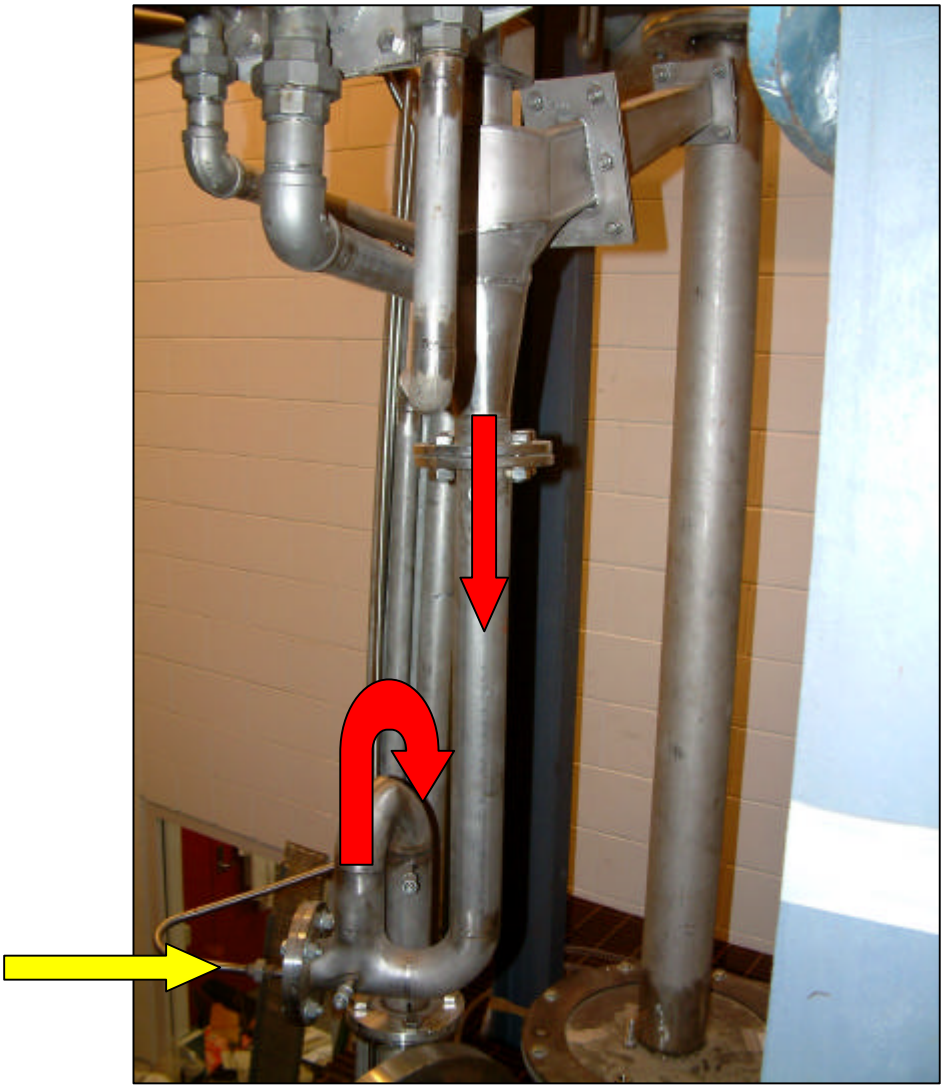

Figure 2-8. The siphon system allows hot bed material (depicted by the red arrows) to transfer from the cyclone (at top of photograph) into the gasification zone but prevents combustion gases flowing back into the exhaust. The U section is fluidised with steam or air fed through the sparge pipe at the left (depicted by the yellow arrow).

Air Supply

Air is supplied to the gasifier with a 50 HP Rootes blower which is also used to supply other equipment in the laboratory. Air is ducted into the laboratory through 80 mm PVC ducts to the valve and rotameter array where it is split into the various supply points. The supply is in three separate zones, gasification column, combustion column and afterburner. The gasification column supply includes the main air supply to the gasification bubbling fluid bed, chute bottom bed material transfer section and siphon top bed material transfer section. These supplies are primarily operated during start up and during bed heat up but are used when the afterburner is not hot enough to generate steam. The air supply for these separates from the main duct then passes through a pneumatic shut off valve operated by the PLC; the supply then splits into the individual lines with separate flow control valves and rotameters on each for flow-rate control. These individual supplies pass through one way valves into the boiler where they are pre-heated with hot combustion gases from the gasification column and the afterburner during heat up, before entering the gasifier at their different positions.

The combustion circulating fluid bed air supply consists of the main fluidising air and also primary and secondary supplies. Like the other air system, the flow passes through a pneumatic shut off valve before splitting into individual supplies. The individual supplies are each controlled with a control valve and rotameter. The combustion air is preheated with heat from the combustion product gases separated from the bed material in the cyclone at the exit to the circulating bed. The main fluidising air and primary air are pre-heated in tube-in-tube heat exchangers; the secondary air is preheated in a shell-and-tube heat exchanger.

The afterburner air supplies are much larger than the others so each has its own shut off control valve. The air is controlled by separate control valves and rotameters on each line. The air to the afterburner is not pre-heated and is directly injected at the three points in the main inlet line, pilot burner and as a dilution supply into the main combustion chamber.

The Afterburner and Steam System

The afterburner combusts product gases from the gasification column to generate steam supplied to the siphon, chute and gasification column. Product gases leave the gasification column, passing through a cyclone to remove small particles of bed material and char entrained in the gas stream. At the top of the cyclone is a sampling port where samples of gas are analysed for composition, and tar and moisture content samples are also taken. The particulate free gas then goes to the afterburner through a duct where it mixes with air and enters the main chamber parallel to the side, creating a turbulent circular flow pattern inside (See Figure 2-9).

The afterburner is constructed from a steel outer section and is lined with Kaowool blanket for protection from high temperatures. The hot surface of the Kaowool is stiffened with special hardener. Product gas flow-rate depends on the wood fuel feed rate and steam feed rate which varies between 15-30 m^3/h. The main afterburner air supply rates range is 100 -170 m^3/h. A pilot burner is situated in the bottom of the afterburner unit operating on low pressure LPG and air from the Rootes blower supply. The pilot operates continuously to ignite the combustible gases and is monitored by a flame rod control system. The temperature of the afterburner is controlled by the dilution air stream which enters the main body of the afterburner just below the boilers. The temperature is maintained below 1000°C by manually adjusting the dilution air.

Figure 2-9. View looking up the afterburner with pilot burner removed. The main duct can be seen entering at the bottom left, product gas flows in the middle duct and air is supplied through the outside. Dilution air is supplied through the duct entering the unit at the top right of this photograph. The boilers can be seen in the middle of the picture.

Steam for the siphon, chute and gasification column is generated in four separate boiler tubes, one for the siphon and one for the chute and two for the gasification bed (the tubes can be seen top centre of Figure 2-9). Valves are used to control water feed rate to the boilers and hence the steam rate. Steam is super heated to 170°C at approximately 10 kPa. The gasification bed uses 6-12 kg/h of steam.

The LPG System

The gasification and combustion beds are both heated to operating temperature by combustion of LPG fuel in the fluidised beds. During operation the bed material temperature in the combustion, bed and hence the whole system, is controlled using LPG to supplement the char being combusted in this bed. The LPG system was designed by Rick Dobbs and installed by Aquaheat Ltd. LPG is supplied to the system from a bank of eight 45 kg LPG cylinders which

is shared with the Fire Engineering Department. Four cylinders feed the system at any one time, so the flow in individual cylinders is not too great to freeze the supply. Operating only half the cylinders at any one time allows for a back up supply and for the cylinder bank to be changed without affecting operations. A regulator reduces the LPG pressure from the cylinder to a line pressure of 20 psi before it flows into the system.

LPG flow into the combustion laboratory, where the gasifier is housed, is controlled by a solenoid valve operated by a key locked switch in the control room. The pilot burners in the both the gasification and the combustion columns and also the afterburner pilot operate on low pressure (3 psi) LPG so these supplies pass through a common regulator. The main LPG supplies to both columns are at line pressure, about 15 psi in the laboratory.

2.2 Construction

Cast Refractory Sections

The gasifier was originally designed to be constructed entirely from cast sections of insulated refractory housed in steel cylinders bolted together. It soon became apparent that it was more favourable to have the more complex sections constructed from stainless steel to allow much easier construction and reduced heat up time without compromising heat losses and other differences from a large scale system. The cast sections consist of four separate layers: the rolled steel outer, calcium silicate board backup insulation, vermiculite insulation and the refractory hot face.

The steel sections are cylindrical in shape with an inside diameter of 390 mm and stand 495 mm tall and are constructed from 4 mm mild steel. Each end of each section has an 8 mm thick flange which has an overhang of 20 mm into the section and has eight evenly spaced 14 mm holes for M12 bolts. The inside of the sections are lined with Calcium Silicate board. On the gasification column the board is 20 mm thick and 50 mm thick in the combustion column which has a smaller internal diameter. The boards are cut to shape and glued to the inner surface of the steel sections with builders' glue (Selley's "No More Nails").

The inside of the calcium silicate board is lined with vermiculite insulation cast on the inside of the calcium silicate board. The vermiculite is mixed with calcium aluminate cement (Electroland brand) in a ratio of seven parts vermiculite to one part cement, which is about one to one on a weight basis. Water is added at a rate of 2–2.5 times the weight of the vermiculite and cement mixture to obtain a consistent mixture which forms balls that hold their shape when compressed. The steel sections have steel rods welded to the inside to hold the vermiculite in place once dried. The inner surface is cast with cardboard tubes. The insulation is then left to dry for 48 hours in a warm dry room. In both the gasification and combustion columns the vermiculite insulation layer is about 40 mm thick.

The hot face refractory lines the inner most surface of the columns where it withstands the abrasion and temperature of the fluidised beds. The refractory was also cast in place with cardboard tubes. The refractory cement is a Sila product, silicast LC 180 which is a 93 % alumina low cement castable product suitable for temperatures up to 1800°C. This product is over specified for the conditions in the gasifier but was gifted to the project. The refractory was mixed into a stiff slurry with water at a rate of 5-7 kg per 100 kg of refractory. The refractory layer in the combustion column is about 50 mm thick giving an internal diameter of 107 mm. The gasification column has a refractory layer about 30 mm thick with an internal diameter of 207 mm. The refractory was vibrated into the mould on a vibrating table for a short period so that water did not separate out from the cement but sufficiently for it to flow into all areas of the cast.

Each of the sections has many ports for thermocouples, pressure tapings, pilot burners and air jets. These were cast into the refractory during construction. Each section contains about six thermocouple ports distributed evenly around the section. The ports all have refractory lining which was moulded during the casting of the refractory. The bottom sections of each column are both quite complex casts containing many ports for chute inlet and the air jets in the combustion column and chute outlet and fuel feed port in the gasification column. Because of these complexities they are cast with Kaowool insulation instead of the vermiculite as it is much easier to work with in confined areas and has more structure in small portions.

The refractory attains full strength after drying and firing processes have been undertaken. Initial drying takes place with the cardboard tubes still in place in a dry environment at room temperature. During this part of the process the refractory attains a solid structure and the

cardboard cast can be removed. The refractory still contains high levels of moisture which, if driven off too quickly, will cause the refractory to crack and lose structural integrity. The sections are then heated individually with a gas burner to over 300°C to drive off water still in the refractory and vermiculite insulation. To help vapour deep inside the insulation to escape to the surface when heated, holes were drilled in the steel shell. During the heating with a gas burner, the temperature of the cast was monitored so it did not heat too quickly. Cardoso et al (2004) show that moisture is driven from the refractory in three phases. The temperature of the cast sections is increased very slowly in these temperature ranges, at a rate of 20°C/h. The final stage of the firing was undertaken once the sections were assembled into columns and were fired with the main LPG burners. The sections were assembled with eight M12 x 60 mm bolts. They were sealed with 10 mm Kaowool blanket gaskets between the refractory to allow for expansion under hot conditions. The sections were sealed gas tight with a high temperature silicon gasket sealant (Ados gasket 260).

Stainless Steel Components

Much of the pipe work and duct work including the cyclones and heat exchangers were constructed from 253 MA grade stainless steel. As it has an operating temperature of up to 1200°C, 253 MA grade stainless is more than adequate for this application. The drawings for these sections of the gasifier were made by Rick Dobbs and they were constructed and welded by Charlston Engineering, Darfield. The working drawings can be found in the ‘Gasifier Operational Guide’. The pipe-work is constructed from standard BSP pipe in sections so it can be easily taken apart. The threaded components are painted with nickel anti-seize compound (ROCOL brand). In large scale plants the low structural strength and high thermal expansion of stainless steel makes it not suitable for construction. Erosion and stress cracking of welds can also be a problem over the long term.

3 Commissioning

This phase involved testing of the burner systems, cold testing the fluid beds and bed materials and developing the procedure for running the plant. Commissioning firstly involved testing the fluid beds, siphon and chute to make sure they worked as designed, then running the plant. This initial cold testing occurred before construction of the plant had finished. Initially the plant was operated at hot conditions without wood fuel and the pilot burners were redesigned and adjusted to achieve consistent operation. Wood fuel was then introduced to find the best method for operating the gasifier under gasification conditions. The procedure developed for operating the gasifier is attached as Appendix C, “Gasifier Operations Guide” and is a work in progress as modifications are continually made to the gasifier to make it operate more satisfactorily.

3.1 Cold Fluid Bed Trials

Once the plant was constructed and the refractory sections were fired to above the critical temperature of 600°C with no bed material inside, the fluidised beds, chute and siphon were tested to make sure they operated as intended. Before the columns were constructed the bed material was tested in a small fluidised bed to find the minimum fluidisation velocity and to see how it behaved under certain conditions. The chute was tested to find the fluidising gas flow-rate required to allow bed material to flow between the two columns. The columns were then reassembled and the siphon was tested by circulating bed material out of the combustion column. The effect of the superficial gas velocity in the combustion column on the circulation rate was measured by increasing the combustion column air and measuring the circulation rate. The top of the gasification column was left open so the bed material transfer could be monitored.

Minimum Fluidisation Velocity

The bed material was initially tested in a separate fluidised bed to find the minimum fluidisation velocity which is important for operating the plant. Air was supplied to a distributor in the bottom of the bed from the CAPE department’s Rootes blower. The air flow-rate is controlled by a valve and rotameter. The flow-rate to the bed was increased incrementally and the pressure drop across the bed was measured for each change in flow as

shown in Figure 3-1. Two different bed materials were tested; both greywacke plasterers' sand with different particle size distributions. The first bed material (sand 1) was screened through a 1000 µm screen only to remove the large particles. The second (sand 2) was screened through a 1000 µm screen to remove large particles then with a 350 µm screen to remove the smaller particles.

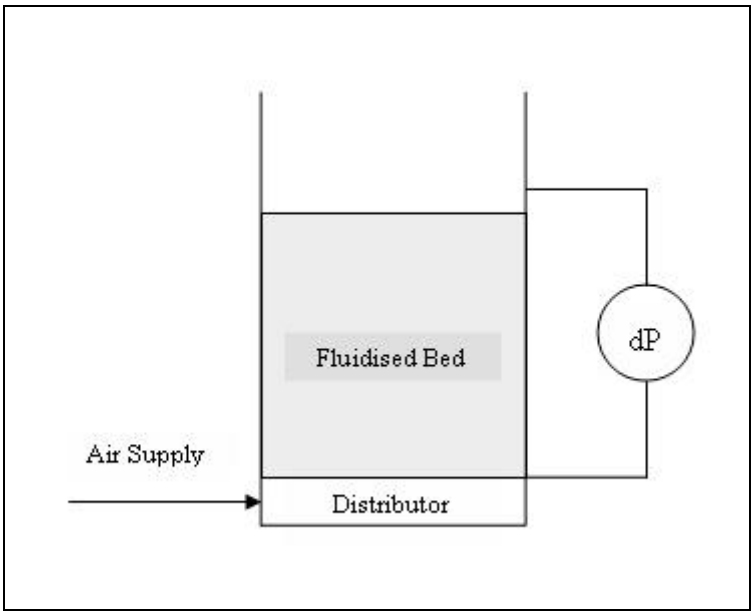

Figure 3-1. Testing set to measure minimum fluidisation velocity of the CAPE gasifier bed material.

Figure 3-2 shows the results of the experiment to test the minimum fluidisation velocity of the gasifier bed material, plotted as bed pressure drop against superficial gas velocity. From this plot it can be seen that sand 1 with the small particles remain fluidised at a lower velocity of 0.29 m/s. Sand 2 with the small particles removed was fluidised at 0.36 m/s. This implies that, during gasifier operation, the gasification column superficial gas velocity must always be maintained at above 0.36 m/s as the bed material used is screened through 1000 µm or smaller. The effect of increased temperature will increase the minimum fluidising velocity as the decrease in density will reduce the ability to fluidise the bed.

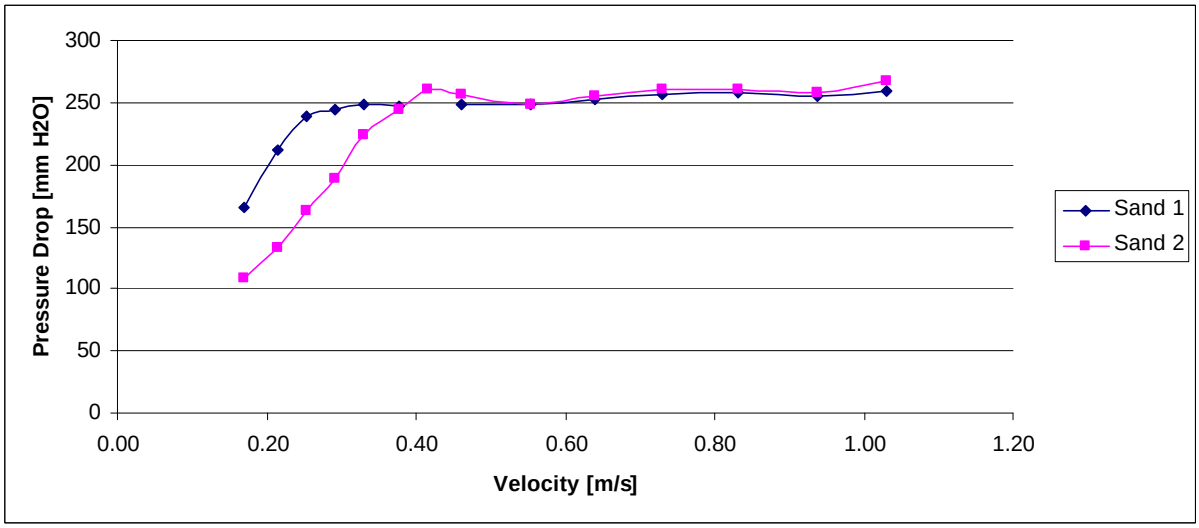

Figure 3-2. Results of testing the minimum fluidisation velocity for the CAPE gasifier bed material.

Chute Testing

These tests were conducted to find the minimum fluidisation velocity required for bed material to transfer between the gasification column and the combustion column. The testing involved adding bed material to the gasification bed and removing it from the combustion bed while adjusting the chute air rate until bed material flowed between the two beds as shown in Figure 3-3. Once the columns were reassembled, the effect of the chute fluidising air rate was tested by keeping all the other flow-rates constant and adjusting the chute air rate whilst measuring the circulation rate with a bucket and stop watch. The bed material used was the standard mix of plasterers' sand (Greywacke river sand, purchased from Placemakers) used in the gasifier. This was screened to remove granules greater than 1000 µm.

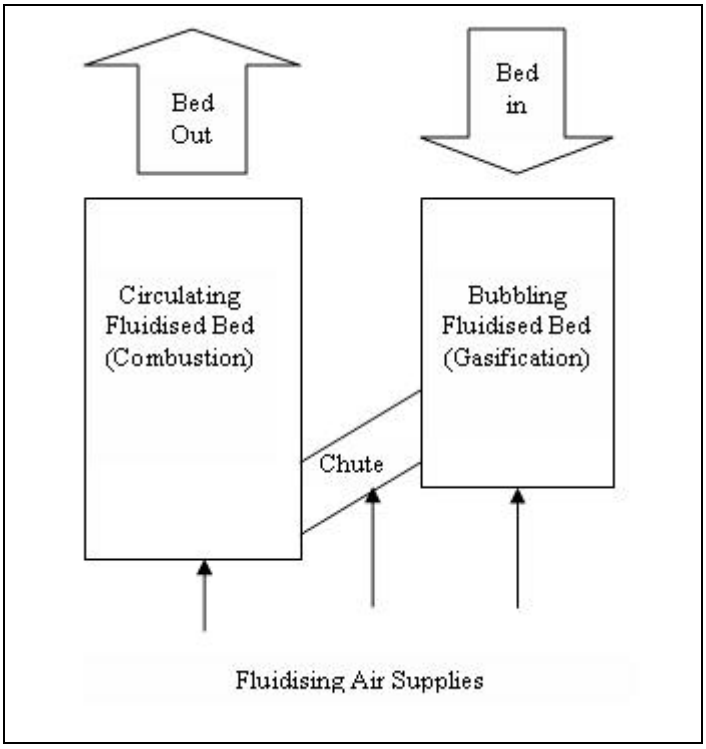

Figure 3-3. Testing the fluidisation velocity of the bed material transfer chute involved bolting the bottom sections of each column together and fluidising them. Bed material was added to the gasification side and removed from the combustion column.

The initial cold test of the chute in the disassembled bottom column sections showed that the chute required a fluidising air flow-rate of between 40 and 50 L/min to initially start the bed material flowing between the columns. However, once the flow had started, the fluidising air flow-rate could be reduced while maintaining the circulation of bed material but it was visibly obvious that the rate reduced as the flow-rate was reduced until circulation stopped when the flow-rate dropped below 20 L/min. To prove that the circulation rate was influenced by the chute fluidising air rate, the circulation rate was measured for variations in the chute fluidising air rate while keeping all other flow-rates constant. The flow-rates used in this experiment are shown in Table 3-1. Figure 3-4 shows that the bed material circulation rate was influenced by the chute fluidising air rate.

Table 3-1. Air supply flow-rates during testing of bed material circulation rates' dependence on the chute fluidising air rate.

Air Supply Point	Flow-rate (L/min)
Combustion column fluidising air	407
Combustion column primary air supply	1710
Combustion column secondary air supply	621
Gasification column fluidising air	496
Siphon fluidising air	113
Chute fluidising air flow-rate	Variable

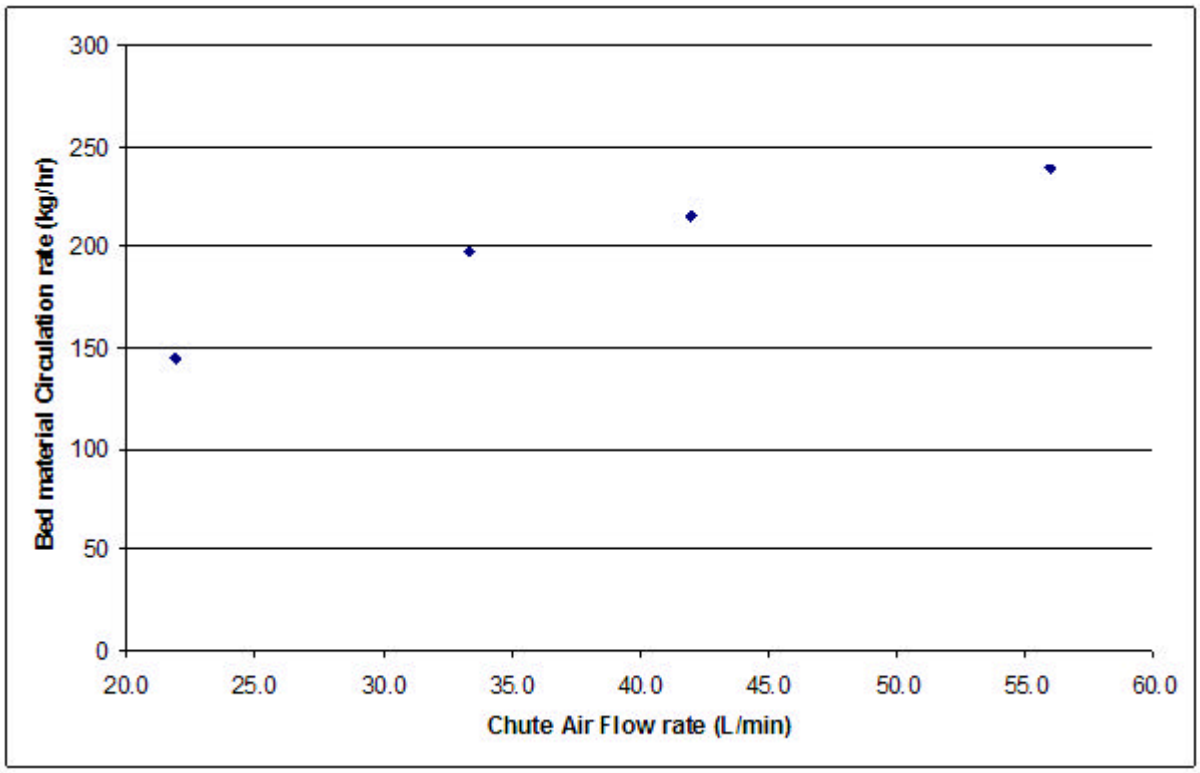

Figure 3-4. Plot of the circulation rate against chute fluidising air rate with all other air flows kept constant.

Combustion Column Velocity vs. Circulation Rate

The bed material circulation rate is very important in the heat transfer from the combustion column to the gasification column. Other than the temperature of the bed material, it is the only way to control the heat transferred. This is important because it illustrates that if the bed material circulation rate can be increased and controlled and then the gasification temperature can be controlled more effectively.

To measure the circulation rate of bed material between the combustion column and the gasification column, the top of the gasification column was removed so samples could be collected. The bed material dropping out of the siphon onto the top of the bed was collected in a 1 L beaker. The time to collect the material was measured and the weight of bed material recorded. The superficial gas velocity in the combustion column was controlled with the air flowrate to the column. The velocity was varied with the air flow-rate between 0.38-0.62 m/s.

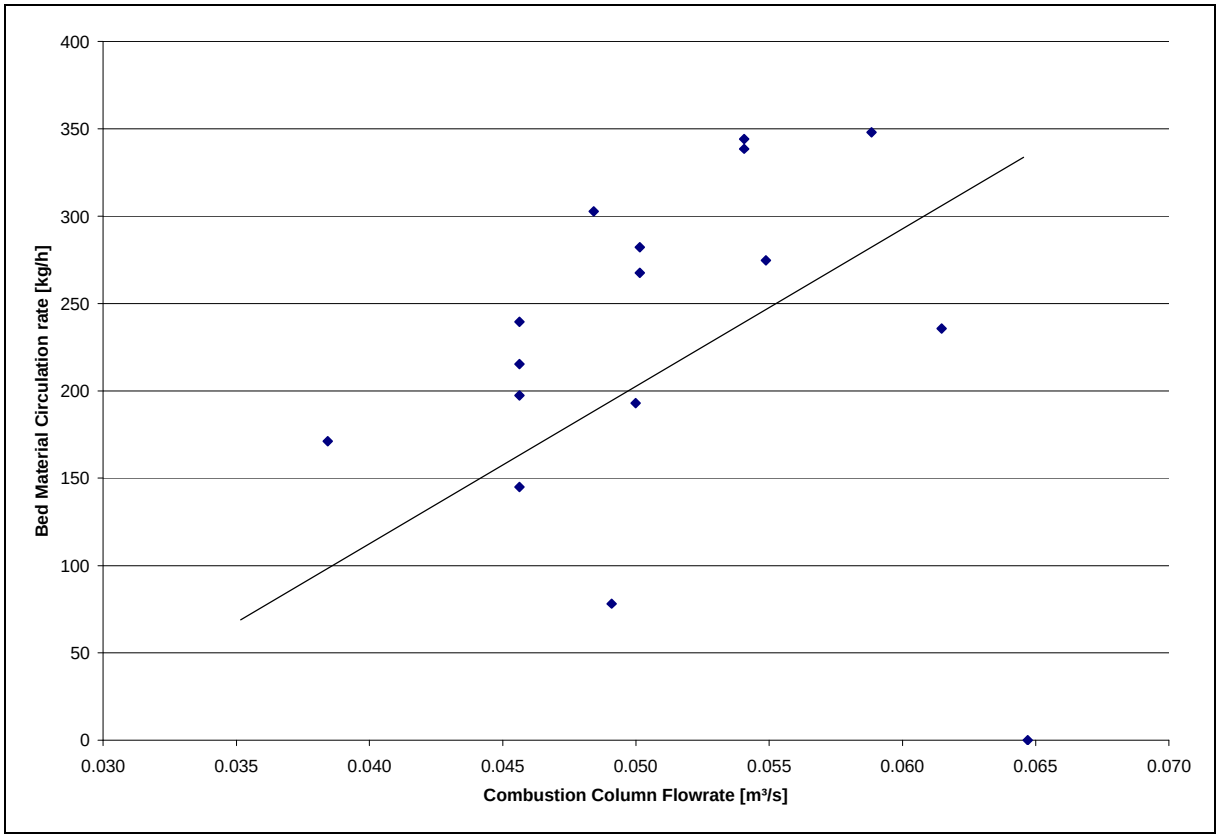

Figure 3-5. Plot of combustion column superficial gas flow-rate against bed material circulation rate.

The combustion column superficial gas velocity influences the bed material circulation rate, increasing it as the column gas velocity is increased. Figure 3-5 shows the upward trend in bed material circulation rate as column superficial gas velocity is increased. Although this figure provides values for the circulation rate and how it is influenced by the velocity, it does not provide a realistic indication of the columns behaviour at high temperatures like those of operation. Hence we can not use the column's velocity to set the circulation rate under operating conditions. Another feature shown during this testing was the tendency of the combustion gases to blow straight through the siphon when the air rates are high and there is a

large back pressure at the top of the stand pipe. When operating under gasification conditions this is unsatisfactory as combustion gases getting into the raw product gas will dilute it and reduce the calorific value, defeating the purpose of having separate gasification and combustion zones. Further investigation and understanding of the hydrodynamics of the bed material circulation is required to reduce product gas dilution. The development of the siphon is critical to maximising the bed material circulation rate whilst minimising combustion gases diluting the raw product gas.

3.2 Pilot Burner Testing

After initial testing of different over-fire pilot burner prototypes similar designs were adopted for both the gasification and combustion columns. The pilot burner design currently operated in the gasification column and initially operated in the combustion column circulating fluidised bed is shown in Figure 3-6. These burners operate on low pressure LPG controlled with a regulator, fed into the pilot burner through a 5 mm tube. The burner assembly shown in Figure 3-6 is housed inside a 25 mm tube protecting it from the bed material which can extinguish the flame or prevent the flame rod from detecting the flame. Compressed air is fed to the burner from the main department compressor and the flow-rate is controlled with a regulator. The compressed air is also fed via a 5 mm tube with multiple holes at various points along the burner which create swirl. The flame is ignited with an electric arc sparking between an ignition rod and the LPG supply tube. The igniters start 20 s before the LPG and air solenoid supply valves open and will start again when a flame is no longer detected. If a flame is not detected for more than 2 s the LPG and air supplies are cut. If the burners do not restart on a second attempt, the controller locks out and needs to be restarted manually.

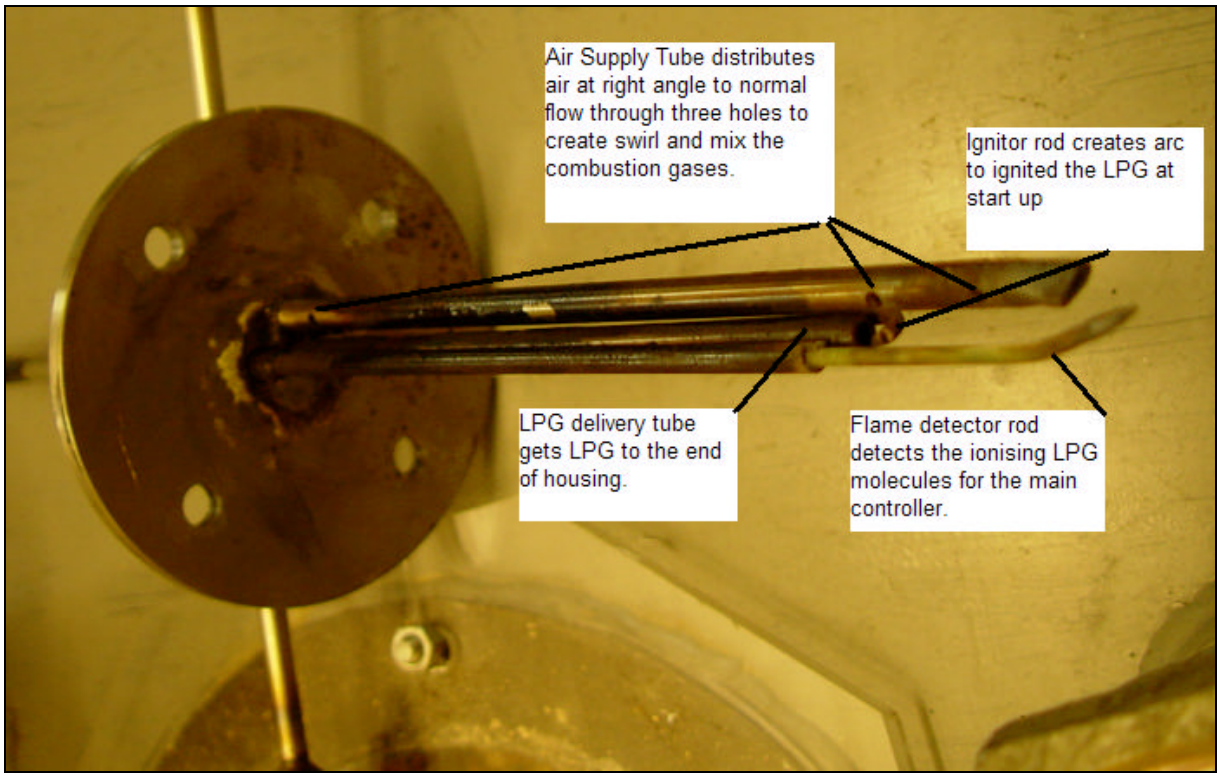

Figure 3-6. This burner configuration is housed inside a 25 mm tube which protects it from bed material which can interfere with the flame and detector. Air is delivered at right angles to the main flow of gas in the tube to create swirl and mix the LPG with the air making it harder to extinguish the flame.

These burners were tested in both beds and operated successfully with high LPG and air rates. The high air and LPG rates give a very stable flame and high velocity in the outer tube prevents bed material getting into the assembly in the combustion column which is a circulating fluidised bed. Bed material that gets into the assembly can extinguish the flame and interfere with the flame detectors. Unfortunately the high LPG and air rates also generate very high temperatures which caused bed material to fuse with ash forming a cone over the end of the burner where it enters the combustion column as shown in Figure 3-7. The hard fused material often broke off and caused blockages in the restricted areas of the siphon and chute reducing bed material circulation and in one case caused hot bed material to be forced through the air preheating heat exchangers. To reduce the temperature of the flame entering the combustion column without reducing the velocity in the tube, a secondary air supply was added to the outside of the assembly housing tube. This system worked very well and dramatically reduced the problem on short runs.

Figure 3-7. Cone of fused ash and bed material formed over the end of the combustion column pilot burner due to the high temperatures of the flame.

However, the burners only operated well on short gasification runs as the secondary supply became blocked when operated for 5-6 hours which allowed the temperature of the burners to rise. The final straw came when the burner blocked and continued to burn inside the housing tube, eventually melting it. The burner got so hot, the refractory cast which houses the pilot burner failed and cracked, exposing the insulation and steel outer to very high temperatures, which burned off the paint. It was clear that a new burner configuration was required. Three different set ups were tested in a high velocity bubbling bed and one in atmospheric conditions outside the bed.

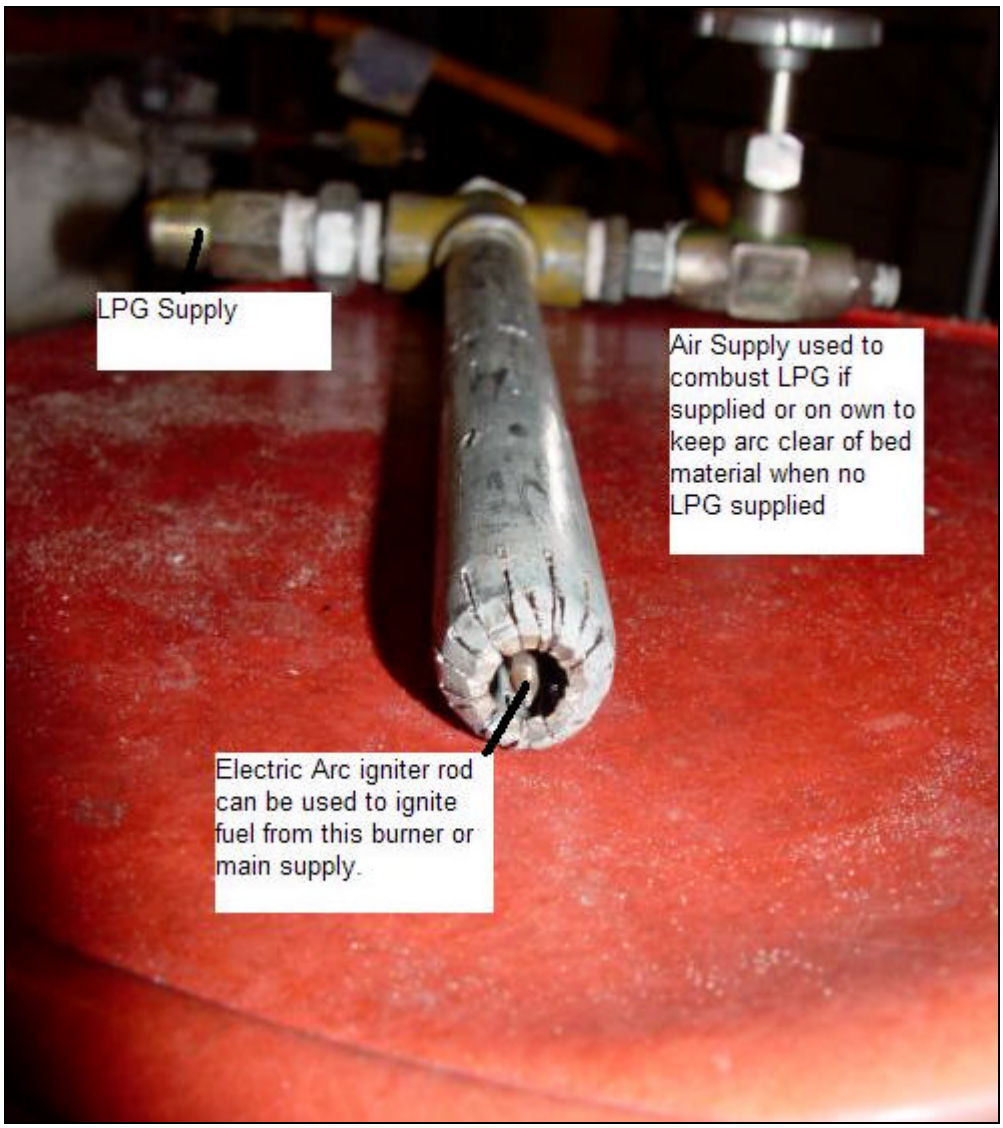

Figure 3-8. This trial burner operates with a continuous electric arc. LPG can be fed to make a small flame to ignite the main supply or arc alone can be used as the sole ignition source. Compressed air can also be fed to combust LPG or just to keep the arc clear of bed material when no LPG is fed.

The first ignition system tested was an electric arc inserted into the bed where the pilot burners are normally situated. Air is fed down the tube to keep the arc clear of bed material which could earth the flame rod and stop the arc. The end of the tube is tapered to encourage the arc to only form at the tip. The ignition rod is insulated along ¾ of its length and has a pointed end to encourage the arc to only form at the end. LPG can also be fed along the tube to create an unstable flame which helps ignition of the main burner.

The electric arc configuration successfully ignited the LPG from the main burner in the bottom of the bed; however, there were three issues that made it unsuitable for permanent use. The first problem is the flame control system and sections would need modification to allow for a photo-cell to be used for detecting a flame. Currently there are no ports in either column to allow for a photo-cell to be installed. The controller is configured so photo-cells can be installed but the photo-cells will need to be purchased. Secondly during operation for long periods of time bed material began to fuse between the rod and the surface of the housing tube which eventually lead to a short circuit and no arc. This problem could easily occur during operation; therefore this ignition source is not appropriate for our application. Thirdly but insignificantly in light of the other issues, the combustion produced by ignition with an electric arc was not very stable because there is not perfect mixing of the LPG and air gas phases in a fluidised bed. Unlike a flame, the arc only provides an ignition source in a small part of the column.

A third type of burner was tested briefly outside the columns to see if it was suitable for further investigation. This burner consists of a compressed air and low pressure LPG supply as with the other burners but the gases are mixed in a combustion chamber where they are ignited by a spark plug which runs continuously. The gases ignite in an explosive manner throwing a flame down a small diameter tube. To get the flame to burn all the way down the tube the mix of LPG and air is important; if the flame does not reach the end of the combustion tube it will not ignite the main burner. The small tube diameter of this configuration means there is very high velocity in the tube meaning bed material cannot get into the burner. Since the ignition arc is in a combustion chamber at the top of the tube, well protected from bed material the flame is very hard to extinguish. Unfortunately this type of burner can not use a flame rod for detection because of the intermittent nature of the explosive combustion. Because of this a photo-cell would be required for detection and modifications to

the columns would also be necessary to house a new detection system. This certainly is an option that should be kept in mind if and when issues arise again in the future.

The fourth burner tested is a variation of the original burner described in the first paragraph of this section. Low pressure LPG is supplied through a tube at very low rates compared with the old burners. This is ignited by a continuous electric arc which operates between a rod and the LPG supply tube. Compressed air is supplied directly into the housing at the base of the burner. The flame in this burner can be detected with a flame rod and no alterations to the columns or burner controller is required. The housing tube has been extended on this model so it enters the column past the coned section so it sits flush with the inner surface of the column.

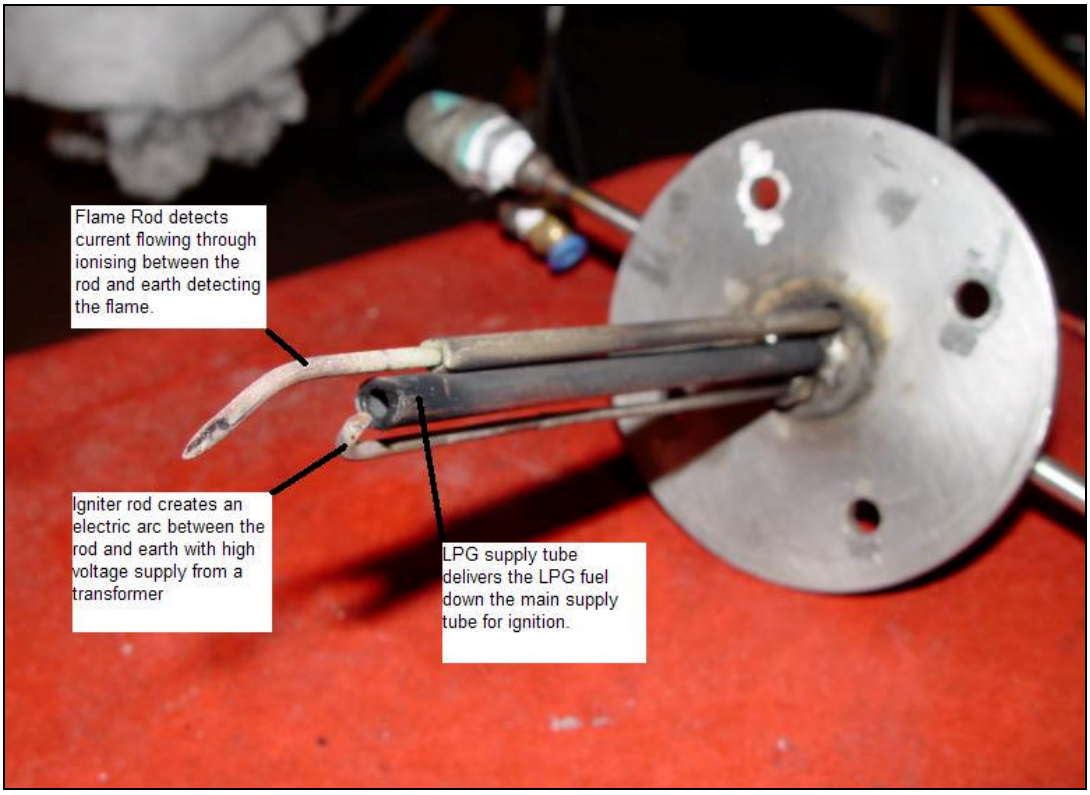

Figure 3-9. Trial LPG burner set up trialed in the CFB. This burner operates with low LPG and air rates to keep the temperature low and with a continuous arc to prevent the flame being snuffed out by the bed material. No air swirl tube is used in this burner to help mixing of air and LPG, air is fed directly from the back of the burner.

As with the other burners this configuration was tested outside the bed in the open air to check a stable flame could be produced then tested inside the fluidised bed. The burner operated with a stable flame and gave a satisfactory signal to the burner management controller. Operating in the fluid bed the burner provided a satisfactory signal of 40 to 50 µA

and was not extinguished by the action of the bed material in the violent environment of the circulating bed. Based on these findings this burner configuration was used when the damaged refractory was rebuilt.

3.3 HAZOP Safety Review

The commissioning process involved a rigorous safety process which included a hazard and operability (HAZOP) review. On the 20th and 21st of December 2005 the CAPE gasification group was involved in the HAZOP review; the meeting was chaired by Eric Scharpf from Delta S Consulting. This procedure was a major part of the commissioning phase so the HAZOP document is attached as Appendix E. The procedure involves breaking the process down into nodes then applying key words to each of the nodes to identify possible hazards then working through each to decide the likelihood of the hazard taking place. The CAPE gasifier was broken down into three nodes: the feed system and gasification column, the combustion column and air pre heating heat exchangers, and the afterburner and steam system. For each node the parameters, such as flows of different materials, temperatures and power and air failure were identified and key words applied to assess the cause and effect of each hazard. As the main operator the author's job during this process was to speak of experiences gained operating the plant during the commissioning. Documentation of the gasifier used in the review was put together by Eric Scharpf, Rick Dobbs and the author. The documents used in the review included:

1. Piping and Inst. Diagram (Flowdiagram2.pdf undated)
2. Risk Analysis Update (undated)
3. Safety Briefing Sheet (undated)
4. Training Slide Handouts (14 Dec 2005)
5. HAZOP worksheet example (14 Dec 2005)

As a result of this review many changes were made to the operating procedures and more testing and calculations were required to check assumptions made in the design of the gasifier and extraction system. The flame supervision system was identified as a key safety feature of the process on which many other safety features were dependant. An evaluation of its effectiveness was commissioned to decide on further action required. Evaluation of the flame management system includes a review of operating without the pilot burners at elevated temperatures. Emergency lighting in the laboratory to operate during a power failure was identified as another major concern. To identify refractory failures heat sensitive paint on the outer steel casing should be considered for the insulated refractory sections. Operation of the electric arc and other equipment in the same laboratory is to be considered separately of the study to have a whole department policy. Procedures for handling the inert gases and flammable gases used in this process also require special treatment.

4 Experimental and Analysis

The experimental and analysis involves measurement of the gasifier product gas composition and the composition and quantity of the tars in it.

4.1 Product Gas Analysis

Product Gas Conditioning Train

Dirty product gases from the gasification column contain tars and particulate matter which needs to be removed before the composition can be measured using gas chromatography (GC). The gas cleaning train removes the tars and particulates in a series of particulate filters and wash bottles based on the European Union Guideline for Sampling and Analysis of Tar and Particles in Biomass Producer Gases (Simell, Stahlberg et al. 2000; Neeft, Knoef et al. 2002). Tars are sampled from the train using the solid phase extraction method (Brage, Yu et al. 1997). Amino phase extraction columns are used to collect the tar material and it is analysed using high pressure liquid chromatography (HPLC). The method was developed by Sean Devenish, a post doctoral graduate from the University of Canterbury's Chemistry Department.

The inlet to the product gas clean-up and analysis train is situated at the exit of the product gas cyclone. The product gases then pass through a particulate filter to remove entrained bed material and char from gasification column. The filter temperature is maintained above 350°C to prevent tars condensing on the surfaces, provision was made to trace heat the filter but the hot producer gas with insulation on the tube was sufficient to maintain the temperature. Quartz thimble filters (26 mm ID x 60 mm long) were purchased from Scientific Supplies (product number; QF-20).

Directly after the particulate filter is a T section where amino solid phase absorption columns are connected into the line to take gas samples to analyse tars. The details for the tar analysis are explained in detail in section 4.2. The gases then pass through six wash bottles connected in series to remove tars, any remaining particulates and water vapour. The micro size of the tubes and columns in the Micro GC mean particulate matter and tars need to be removed. The water vapour must also be removed as it gets held up in the Micro GC column pores and has a long elution time. The wash bottles are connected in series with the temperature in the first

four being maintained at approximately 15°C by a water bath with continuous flow through it. The wash bottles in the water bath are half filled with clean industrial grade iso-propanol (IPA) in which the tar compounds are almost totally soluble (Neeft, Knoef et al. 2002). The final two wash bottles are kept at very low temperature by keeping them in a mixture of acetone and dry ice. These wash bottles are filled with glass wool to increase turbulence and condense out any remaining water, tar and solvent.

The gas samples are sucked through the system with a diaphragm pump which is situated after the wash bottles as it can only operate with clean dry gas. The pump is a Capex V2 12 V DC diaphragm pump connected with power supply and a variable voltage output for varying the pump speed. The pump can move up to 3 l/min at 2 bar_g but in this application is restricted to about 1.5 l/min or 90 l/h at 0.7 bar_g. The flow-rate of the gas is measured using a rotameter; there is provision for a gas volume totalizer but it is not installed at this stage. The gas totalizer would allow the wash bottles to be used for quantitative tar analysis if installed. The Micro GC has an entrained liquid and particulate genie installed on the front end for protection. The genie requires an inlet pressure of 10 psi so the system down stream of the pump is kept at positive pressure. The pressure is monitored manually and adjusted using a needle valve down stream of the Micro GC supply point.

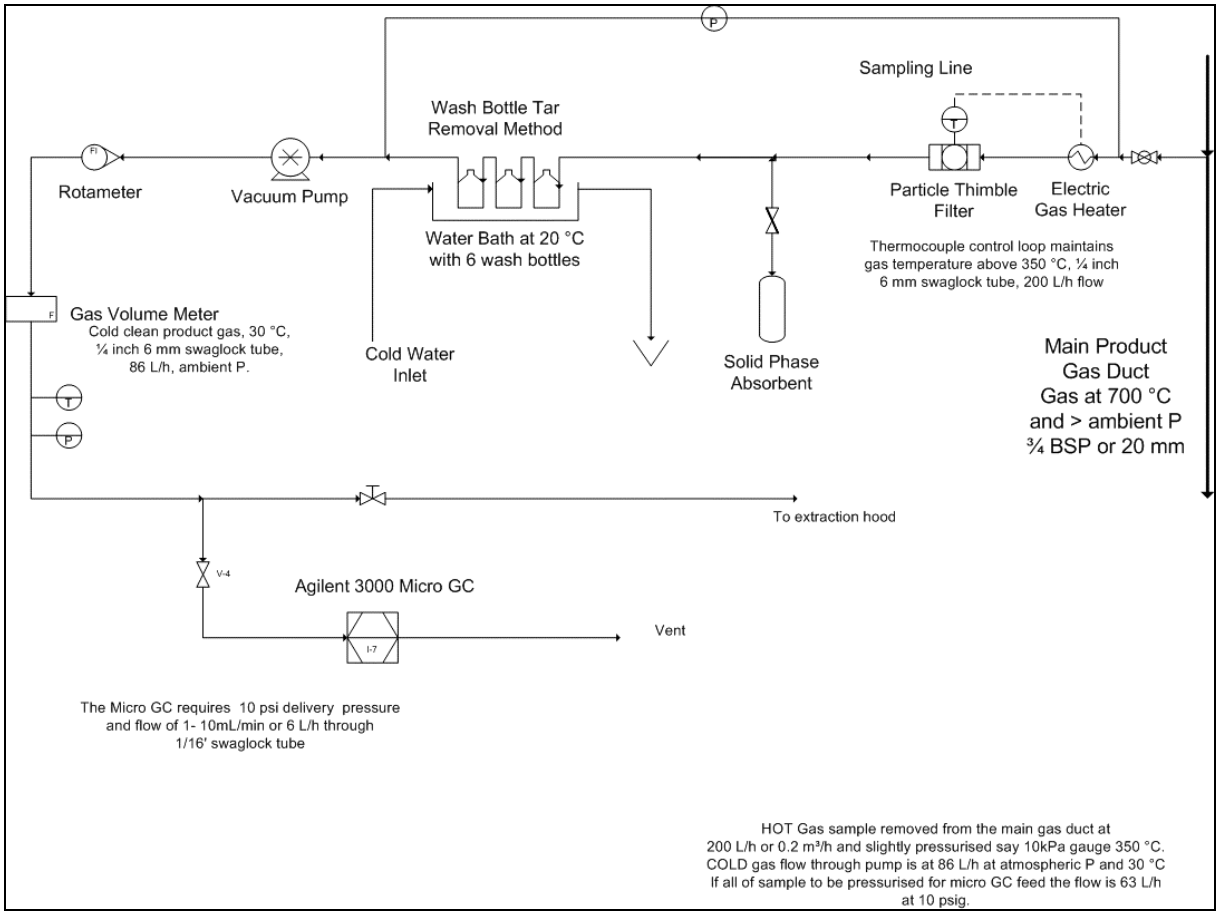

Figure 4-1. Product gas cleaning and gas sampling train.

Micro Gas Chromatograph set up and Operation

In April 2005 an Agilent 3000A Micro Gas Chromatograph (GC) was purchased from Analytical Technologies (a division of Biolab), to analyse samples of product gas from the biomass gasifier. Colin Welkin is the representative who recommended the system purchased and provided much needed support and training after installation. The chromatograph is set up with two channels fitted with Molecular sieve and Plot Q columns for separating the different components that make up the product gas. An on board gas cylinder allows carrier gas for one channel to be stored in the machine. The channel configurations are summarised in Table 4-1. The Plot Q column is fitted with a back flush valve which opens after the analytes of interest have passed through the pre-column to flush unwanted analytes back through the injector. This keeps the analytical column clean and allows it to hold a calibration longer. Both columns are fitted with single filament thermal conductivity detectors which measure the thermal conductivity of the analytes relative to the carrier gas. Both columns also have heated inlets.

Table 4-1. Micro GC configuration.

Channel	**Channel A**	**Channel B**
Column	Molecular Sieve	Plot Q
Carrier Gas	Argon	Helium
Injector	Back-flush	Fixed volume
Detector	Single Filament Thermal Conductivity	Single Filament Thermal Conductivity

Instrument grade helium and argon carrier gases were purchased from BOC gases. The compositions of these are shown below in Table 4-2.

Table 4-2. Composition of carrier gases as supplied by the supplier, BOC, on delivery.

Helium		**Argon**	
Component	**Volume %**	**Component**	**Volume %**
Helium	>99.99	Argon	<99.99
Argon	<5 ppm	Carbon Dioxide	<1 ppm
Carbon Dioxide	<5 ppm	Carbon Monoxide	< 0.5 ppm
Carbon Monoxide	<5 ppm	Hydrocarbons (as Ethane)	< 0.5 ppm
Hydrocarbons (as Ethane)	<1 ppm	Hydrocarbons (as Methane)	< 0.5 ppm
Hydrocarbons (as Methane)	<1 ppm	Moisture	< 5 ppm
Moisture	<10 ppm	Nitrogen	< 5 ppm
Nitrogen	<50 ppm	Oxygen	< 5 ppm
Oxygen	<10 ppm		

The operating conditions of the micro GC channels are shown in Table 4-3; initial values for these were taken from Feeney et al (2003). The operating conditions were changed by trial and error, until satisfactory separation for the gas composition measured in the FICFB gasifier system was achieved. The chromatograph were set by first thermally cleaning the columns at 290°C for at least one hour until a uniform flat base line was obtained, baking out unwanted analytes captured during previous operation. The chromatograph was calibrated before each use with 3 separate samples of the same beta standard calibration gas. The GC was calibrated using a beta standard specification gas mixture made up by BOC gases. The beta standards are calibrated to ±0.5 vol.% uncertainty for the most inaccurate component. The composition and uncertainty of the calibration gas is shown in Table 4-4

Table 4-3. Micro GC operating conditions

Condition	Channel A	Channel B
Sample Inlet Temperature	95°C	95°C
Injector Temperature	95°C	55°C
Column Temperature	110°C	60°C
Sampling Time	15 s	15 s
Injector Time	10 ms	15 ms
Run Time	180 s	240 s
Column Pressure	207 kPa	138 kPa
Backflush Time	15 s	0

Table 4-4. Composition of beta standard used to calibrate chromatograph.

Component	Volume %	Uncertainty (± Vol. %)
Hydrogen	44	2
Carbon Dioxide	27	1
Carbon Monoxide	11.0	0.4
Methane	10.5	0.4
Nitrogen	3.1	0.1
Ethane	0.55	0.02
Ethene	3.7	0.1

Moisture Content Measurements

Product gas moisture content was assessed by taking samples of the gas stream at a port above the cyclone on the exit of the gasification column. The moisture is condensed out of the gas stream in a cold trap. A short section of stainless steel tube insulated with ceramic fibre leads the gas into a 100 mL cold trap filled with glass wool. The cold trap sits inside a thermos filled with a mixture of acetone and solid carbon dioxide (ratio of 1:1 by weight) which has a temperature of -77°C (Merck 1986). The sample is pulled through the cold trap by emptying five litres of water from a glass vessel. The cold trap and upstream pipe work are weighed then heated at 105°C in an oven for 20 min to drive off the moisture. They are weighed again to find the mass of moisture. The condensed tars are then washed out of the cold trap with solvents.

4.2 Tar Analysis

Tar samples are taken using the solid phase extraction method. Tar compounds are absorbed onto amino phase extraction columns. Gas samples, of 100 mL are pulled through the columns, leaving the tars behind on the columns. The small sample size means the method is fast compared with other methods used for tar analysis. The columns are then washed with iso-propanol and dichloro-methane solvents to remove the compounds. The solvent samples

containing tars are then analysed using high performance liquid chromatography (HPLC) to analyse the composition and quantity of the tars. The method detects on average 94 % of the compounds in the tar samples (Brage, Yu et al. 1997).

Solid Phase Extraction Capture Materials

Tars are absorbed on to solid phase extraction (SPE) columns which contain 100 mg of amino-propyl silane bonded silica gel. The columns are commercially available from JT Baker (product number 7088-03). The solvents used are dichloro-methane (DCM) and iso-propanol (IPA), both instrument grade with a purity of 98 % or greater. 1 mL syringes are used for preparing SPE columns, 50 mL syringe for pulling the samples through the columns and Luer lock fittings to connect syringe to column and column to sample line. Needles are required to connect the columns into main sample line.

Solid Phase Extraction Capture Procedure and Sample Preparation

Prior to sampling the columns are pre-treated by flushing them with 0.5 mL of DCM then heating them to 100°C for 5 minutes in an oven. Once the columns are pre-treated, a covered needle is attached to one end and a rubber stopper inserted in the other to prevent contamination in the atmosphere until the column is used. A Luer lock needle is fitted to the end of the SPE column so it can be inserted into the 1/16th Swaglok tube fitting, in the sample line. A 50 mL syringe connected to a Luer lock 3-way valve is inserted in the other end of the SPE column to pull a 100 mL sample through the column. Samples are taken by inserting the needle into the sample line and pulling a sample through the column with the attached 50 mL syringe. To get the 100 mL sample required, the syringe is emptied using the 3-way valve and a second sample is pulled through the column with the syringe. The needle is immediately covered and the end sealed with a rubber stopper. The needle is left attached so that any tars that condense inside it during sampling can be analysed giving a more accurate sample. The samples are then analysed the following day using the HPLC method developed by Devenish (2006).

Tar Sample Analysis

The high performance liquid chromatography machine used to analyse the samples is a Hewlett Packard 1100 Series, fitted with an Alltech Adsorbosphere C18 5 µm 250 mm x 4.6mm column. There is an interim report written by Sean Devenish about the development of the HPLC method and identifying the compounds available in the information available on the University of Canterbury engineering network. Infra-red (IR) spectroscopy was used to identified from the compounds in the tar from a standard mixture and the HPLC chromatogram are attached as Appendix F.

5 FICFB Gasifier Results and Discussion

5.1 Gasification Operation and Characterisation

Background

The FICFB dual bed gasification system has been predominantly developed at the University of Vienna in Austria. Initially the concept was used for the gasification of coal in the 1980s but testing of the system with biomass began in the 1990s. A 100 kW FICFB gasifier on which the CAPE gasifier was based (described in detail by Hofbauer, Veronik et al.(1997)) was used to make measurements required for the scale up design of a 8 MWth commercial scale unit operated to produce heat and power for the local community in Gussing, Austria (Hofbauer, Reinhard et al. 2002).

The differences between the CAPE and Vienna gasifiers are in the fuel feed position, top and bottom bed material transfer systems respectively, and the type of bed material used. The Vienna FICFB gasifier feeds biomass fuel into the side of the bubbling fluid bed gasification column as opposed to it dropping on the top of the bed as in the CAPE system. The siphon which transfers bed material between the top of the combustion circulating fluidised bed and the gasification bed feeds bed solids material to the gasification column just above the bed level at the Vienna plant as opposed to above the freeboard on the CAPE gasifier. The combustion and gasification columns are physically further apart on the CAPE plant due to the insulated refractory construction so the chute, which transfers bed material and char between the bottoms of the gasification column and the combustion column, is longer. The final difference is the type of bed material used. The Vienna gasifier a mixture of quartz sand and nickel treated olivine as opposed to the greywacke sand used in the CAPE gasifier. Extensive experimentation has been undertaken on the 100 kW unit at the University of Vienna to characterise it's operation under a range of conditions (the website www.ficfb.at has extensive results (Rauch (2006b, 30th June 2006).). The extensive research effort by the Austrian consortium with this type of reactor makes their system the ideal model for operating the FICFB gasifier at the University of Canterbury and provides a good comparison for progress.

Vienna FICFB Gasifier Results and Discussion

The Vienna work has focused on optimising the operating conditions to get a high calorific value gas with low tar content. The parameters used to optimise these parameters include the gasification bed temperature which is controlled with the bed material circulation rate and additional fuel added to the combustion column. The bed material circulation rate is controlled using the air flow-rate in the combustion column which is a circulating fluidised bed (Loffler, Kaiser et al. 2003). The ratio of fuel to oxidant, which is steam in this case has also been studied for its effect on the product gas quality and composition (Hofbauer and Rauch 2000). The influence of gasification bed temperature on the product gas composition for the Vienna FICFB gasifier is shown in Figure 5-1. The nitrogen content of the product gas is not shown in Figure 5-1 or Figure 5-2 since it is less than 5 % for all the cases and hence insignificant. The nitrogen comes into the system in the void space of the fuel and the transfer chute and siphon completely seal not allowing combustion gases to mix with the product gas. This is not so important for this study but is very important for modelling the system. Heavier hydrocarbon gases such as ethene and ethane are not shown here as they are less than 3 %. Typically the ethene composition is of the order of 2.5 vol% and ethane, 0.5 vol%, both reducing at higher temperatures and steam to fuel ratios (Rauch (2006, 30th June 2006).).

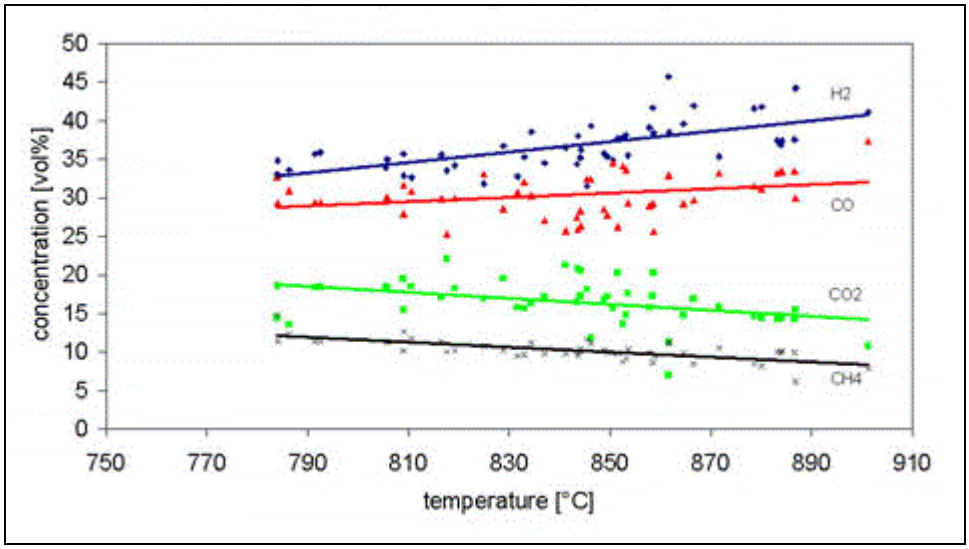

Figure 5-1. Variation of dry product gas composition with gasification bed temperature for a fixed steam to fuel ratio of 0.5.

The hydrogen content of the product gas increases with temperature due to cracking of the heavier hydrocarbons such as tar, ethane, ethane and methane which releases elemental hydrogen. The char-steam shift reactions are also encouraged. The carbon monoxide

concentration is also increased with temperature as increased levels of solid carbon at higher temperatures react with steam in Equation 4, the char-steam shift reaction. Methane is reduced as it is thermally cracked at higher temperatures to form more solid carbon and hydrogen gas. Carbon dioxide concentration is also reduced as temperature increases because of an increase in the amount of solid char formed shifting the Boudouard reaction (Equation 1) equilibrium to form more carbon monoxide. The effects of the steam to fuel ratio on the product gas composition is shown in Figure 5-2. The increase in the hydrogen and carbon dioxide contents and decrease in the carbon monoxide content indicates there is an increase in the homogeneous water gas shift reaction taking place. There is little change in the concentration of methane in the product gas over the assessed range.

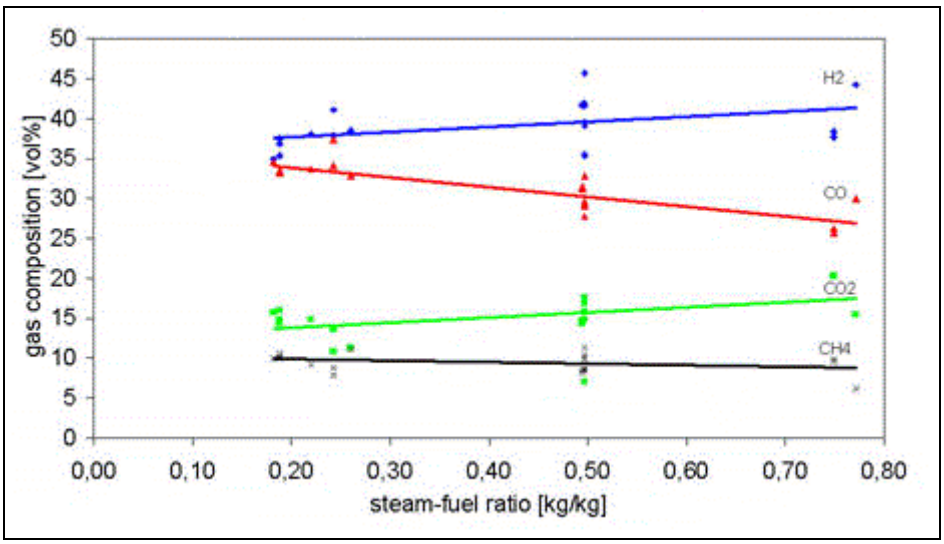

Figure 5-2. Steam to fuel ratio against dry product gas composition with a bed temperature of 850-900°C.

The tar content of the product gases are shown to be dependant on two different factors shown in Figure 5-3, bed temperature and steam to fuel ratio. The tar content follows a decreasing trend for increase in both temperature and steam to fuel ratio.

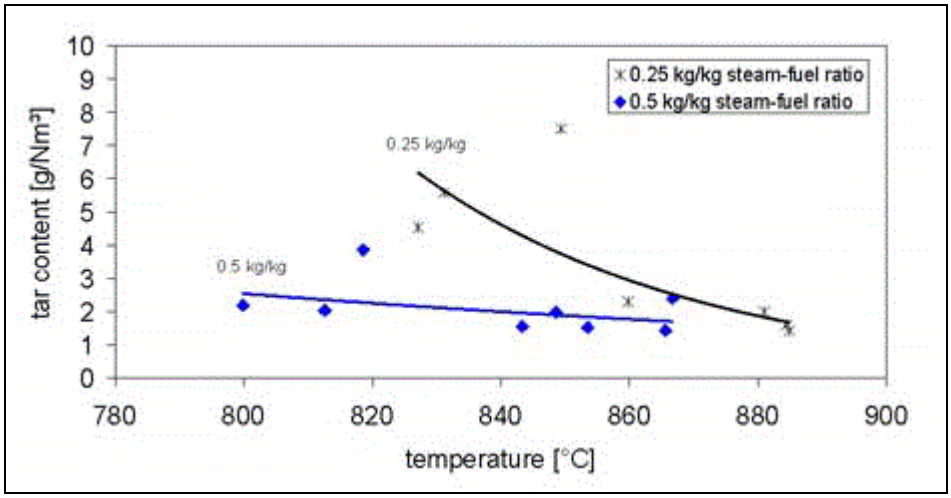

Figure 5-3. Tar content against temperature for two separate steam to biomass ratios.

CAPE Gasifier Results and Discussion

Significantly more data is required for the CAPE gasifier to be able to draw a fair comparison between it and the Vienna gasifier and the Page Macrae gasifier however some design issues need to be resolved before large amounts of data can be acquired at repeatable conditions over long periods of time. The main issues that cause concern while gasifying are the inability to increase the circulation rate to the design specifications meaning low fuels rates are required and only low bed temperatures possible. Segregation of char and bed material in the siphon system means that the combustion column exit cyclone has low separation efficiency for the char. The low separation efficiency of the cyclone results in hot char being blown through the air pre-heaters and being caught in the catcher. The afterburner boilers can not keep up with the steam rates required to provide steam to both the siphon and chute and will require a separate supply so they were fluidised with air for these tests. These issues will be discussed in further detail in the later part of this section. However the data found from these initial tests is presented in Table 5-1. This data shows the steam-biomass ratio for each run along with the gasification column bed temperature and the gases analysed by the micro GC which does not analyse water. The steam to biomass fuel ratios in this table are inaccurate due to a leak in the main boiler down stream of the water feed rate measurement.

The bed material circulation rate is calculated using a heat balance which uses the heat required to produce the product gases at the measured temperature to find heat transferred in the bed material. The material and energy balance is attached to this thesis with an

explanation of its use as Appendix D. The product gas composition is measured using the methods described in previous section, with a micro GC. The gas moisture content measurement method is also described in the same section. The amount of char circulated is calculated with an oxygen balance in the combustion column. The LPG feed rate and oxygen content are measured and the remaining oxygen combusted is assumed to be char circulated from the gasification column. The product gas flow-rate is calculated with a carbon balance on the gasification column, assuming no combustion gases are transferred through the siphon or chute into the product gas stream. A measure of the product gas flow-rate would be better than using a balance as it would increase accuracy. This would also give a better idea of how the product gas flow fluctuates during operation.

Table 5-1.Data collected during operation of the CAPE gasifier, compositions on a dry gas basis as analysed by the micro GC.

Steam-Biomass Ratio	0.48	0.64	0.8	0.48	0.7	0.68	0.68	0.8
Gasifier Bed Temperature [°C]	693	694	696	721	734	750	750	753
H2 [vol. %]	14.2	16.1	18.2	17.2	22.4	19.3	18.0	21.7
CO [vol. %]	26.7	29.0	24.6	27.0	28.9	26.0	25.5	28.4
CO2 [vol. %]	17.4	17.5	19.5	17.6	17.8	16.8	17.6	17.4
CH4 [vol. %]	9.7	11.0	10.2	10.6	12.1	9.2	9.2	11.6
Ethene [vol. %]	2.6	3.1	2.8	2.8	3.4	3.0	2.8	3.5
Ethane [vol. %]	0.6	0.6	0.6	0.6	0.6	0.5	0.5	0.5
N2 [vol. %]	28.7	22.6	24.1	24.2	14.8	25.1	26.5	16.9
Lower heating value (LHV) [kJ/Nm³]	10888	12015	10394	12147	12268	12044	11587	11471

Table 5-2. A typical dry gas composition from the Vienna 100 kW FICFB gasifier (Rauch, (2006, 30th June 2006).) compared with CAPE gasifier composition with N2 adjusted to match Vienna gasifier.

Gas [vol. %]	**Vienna**	**CAPE (N_2 Corrected)**
H_2	37.7	20.2
CO	26.2	36.3
CH_4	9.9	13.8
CO_2	20.3	21.9
Ethene	2.5	3.8
Ethane	0.2	0.8

The lower heating value for the gas produced by the CAPE gasifier ranges from 10394kJ/Nm³ to 12147 kJ/Nm³. These heating values are lower than 12614 kJ/Nm³ for the typical Vienna 100 kW FICFB gasifier heating value, due in the main to the high nitrogen content of the CAPE gasifier product gas. The CAPE gasifier has a nitrogen content of 24.2 to 28.7 vol.%

for steam gasification in the main bed but with air used to fluidise the siphon and chute. When either the siphon or chute is fluidised with steam, the nitrogen concentration falls to between 14 and 22 vol. %. Unfortunately there were some problems with the boiler that were not identified during operation which made the steam rate hard to control and mean not enough is fed to the gasification column. Further investigation after testing was complete showed the boiler had a crack in one of the welds which had led to a leak (See Figure 5-4). The boiler system is required to generate steam at up to 0.25 kg/min or in heat terms, 12.6 kW. Modification to the steam system will be required for a new proposed product gas cleaning train so we are investigating the use of a separate steam supply, using the current boiler setup for super heating. Table 5-3 shows the gasifier operation with air feed to the gasification column compared with steam. The table clearly shows increases in the combustible component of the gas with decrease in nitrogen when the gasification column has a steam feed as designed. By eliminating the nitrogen in the product gas, the calorific value can be further improved.

Figure 5-4. The view of the top of the boilers shows the crack in the weld that led to a steam leak.

The methane, ethane and ethene composition of the product gas is slightly higher in the CAPE gasifier than the Vienna gasifier, when the compositions are compared on a nitrogen free basis (See Table 5-2). This is most likely because of the CAPE gasifier is operating at lower temperatures than the Vienna plant which operates above 800°C and with the top of bed fuel feed. Feeding the fuel on to the top of the bed reduces the amount of thermal cracking taking place because the fuel is not necessarily exposed to the temperature of the bed since in lands on top of it. The fuel begins gasifying as it falls on to the bed and does not need to reach to the temperature of the bed. Lower temperature gasification leads to a high tar and hydrocarbon content in the gas. Table 5-2 also shows the hydrogen content of CAPE product gas is lower and the carbon monoxide content is higher than the Vienna plant. This suggests that the homogeneous gas phase reactions which include the water gas shift reaction (Equation 2) are not taking place to the same extent as in the Vienna Gasifier to produce hydrogen. The most likely cause of this is the low concentration of steam going into the reactor due to the boiler leak. Calculations show the Vienna plant has a product gas steam content of approximately 40 vol.% compared with between 20 vol.% and 33 vol.%. One option to increase the time the fuel and gases are exposed to higher temperatures is to feed fuel in the side of the bed rather than on top of the bubbling fluid bed in the gasification column. Product gases formed will then have to rise through the hot bed material into the freeboard then out of the reactor increasing the exposure time to the hot bed material.

Table 5-3. Comparison of gas compositions for different combinations of air and steam feeds. BFB refers to the fluidising medium of the bubbling fluid bed in the gasification column.

Gas [vol. %]	Air Gasification (low air rate)	Air Gasification (high air rate)	Steam to BFB, air to chute and siphon	Steam to BFB and Chute, air to siphon
Hydrogen	12.2	9.0	18.0	21.7
CO	26.1	24.7	25.5	28.4
CO2	16.6	14.7	17.6	17.4
Methane	8.0	7.7	9.2	11.6
Ethene	2.5	2.4	2.8	3.5
Ethane	0.5	0.4	0.5	0.5
Nitrogen	34.1	41.1	26.5	16.9

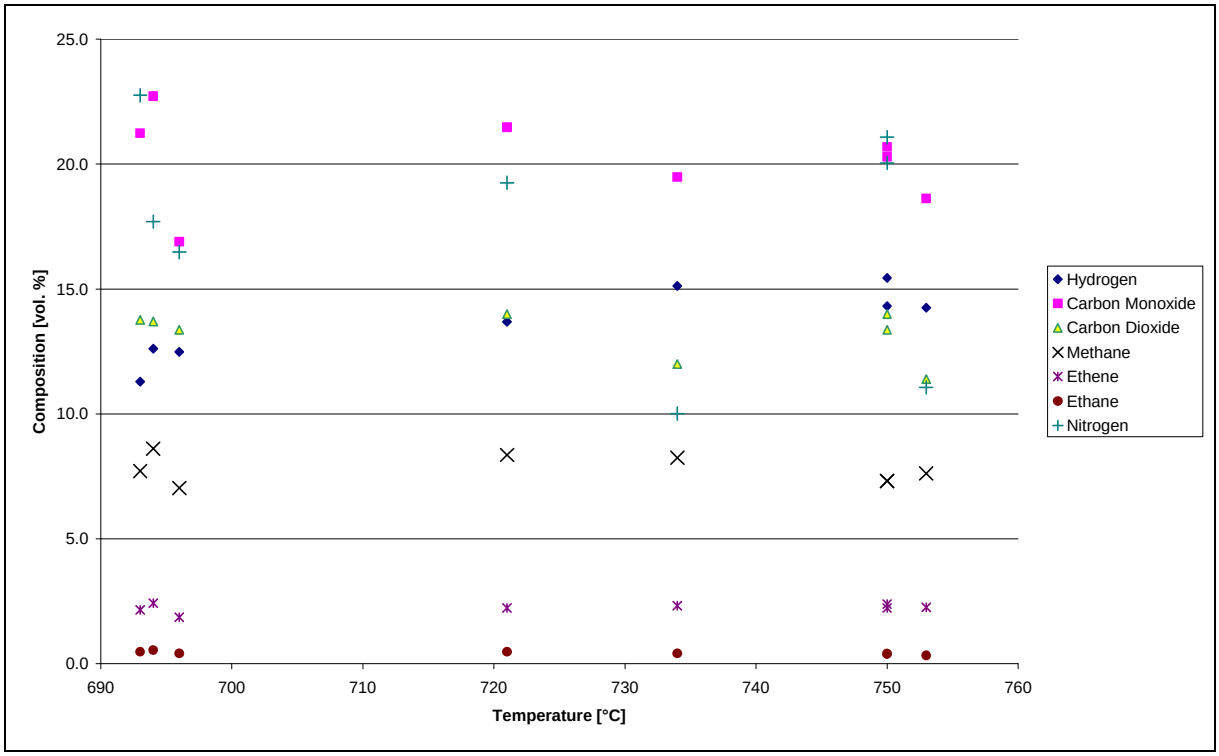

Figure 5-5. Temperature vs. Composition of product gas components.

As temperatures and steam to biomass ratios increase the hydrogen content of the product gas increases as indicated by the blue dot in Figure 5.5 above. The carbon monoxide composition is reduced with both high temperatures, and increased steam to biomass ratio. No trends are apparent in the carbon dioxide, methane, ethane or ethane. However, there is not enough data here to confirm that these trends are significant. Table 5-4 shows the bed material circulation rate compared with the combustion column superficial gas velocity calculated on an ideal gas basis and also the cyclone entrance velocities. The bed material circulation rate calculations are not affected by the leak in the boilers as the moisture content of the product gas is measured and used in the heat balance calculations. The overall plant efficiency for the CAPE gasifier has not been calculated as several parameters are not measured accurately enough to use. The carbon particulates entrained in the product gas stream can not be measured because they are collected in a sealed hopper. The tar in the product gas is not quantified, whichs makes the mass

Table 5-4. Bed material circulation rate with column velocity and cyclone entrance velocity under hot conditions in the circulating fluidised bed (CFB) combustion column

Bed Material Circulation rate [kg/h]	CFB Column superficial gas velocity [m/s]	CFB Cyclone entrance velocity [m/s]	CFB Cyclone entrance velocity (allowing for bed loading) [m/s]
613	5.4	17.1	17.8
580	5.5	17.5	18.1
486	5.0	16.1	16.2
524	5.2	16.5	16.6
535	4.7	15.1	15.6
567	4.9	15.6	16.1
779	5.1	16.4	17.3
790	5.1	16.2	17.0

The design bed material circulation rate is 1200 kg/hr calculated by Rick Dobbs to achieve a gasification bed temperature of 850°C, at a biomass fuel feed rate of 32 kg/hr. This calculation assumes a bed material heat capacity of 1200 J/kg.°C, whereas the bed material used has a heat capacity closer to 800 J/kg.°C (Perry and Green 1997). This means the bed material circulation rate required to achieve the design parameters is actually 1800 kg/hr hence the cyclone, siphon and chute are not designed to meet the transfer rates required. Using other optimised parameters from the Vienna and CAPE gasifiers would require a bed material circulation rate of 1965 kg/h to gasify 32 kg/h of biomass fuel at 850°C. The fuel feed has a lower operating limit of 15 kg/h because the motor and variable frequency (VF) drive over heats below this rate. The bed temperature can not get above 750°C with the current restrictive bed material circulation rate, which according to the mass balance has a maximum rate of 790 kg/h under gasification conditions. Under cold bed testing the rate could not be taken above 450 kg/h without gases being blown through the siphon since this can not be monitored under hot gasification conditions it is possible that combustion gases are blown through the siphon at high circulation rates. The amount of air blown through the siphon can be calculated by fluidising the gasification column with pure nitrogen while measuring the oxygen content of the gas exiting the column. The effects of the bed material circulation rate on temperature at operating conditions (temperature, flow-rates, pressure) has not been quantified in this thesis; however, work completed so far shows that further investigation in this area could be very valuable.

Observations made during operation of the gasifier show that when air flow-rates to the combustion column are increased, bed material is blown through the air pre-heating heat exchangers, not separated from the combustion gas stream. The cyclone design inlet velocity

for best separation of the bed material is 19 m/s calculated by Rick Dobbs and checked by the author. The calculated velocities during successful gasifier operation are shown in Table 5-4 and range from 15.6 to 18.1 m/s when the solids loading is taken into account. These are slightly lower than the ideal design velocity but well within the operating range of a cyclone of 15 to 30 m/s. Below 15 m/s there is not enough velocity to separate the particulates with centrifugal force. Above 30 m/s interactions between the high velocity particles is high and the bounce mechanism is dominant (Grace, Avidan et al. 1997). This shows it is unlikely the inlet velocity which is the cause of bed material not being separated in the cyclone. It is possible that interactions between bed material and char could be the cause of some of the problems and may be that the larger lighter char particles need to be separated in a secondary cyclone or stopped from circulating out of the combustion column with a grate or similar mechanism. Extensive cold testing of the cyclone and siphon region is required to solve this problem. Another possibility is the siphon is not able to keep up with the amount of bed material being separated by the cyclone and builds up in the stand pipe and fills the cyclone. Observing the circulation with a transparent stand pipe or window would be very helpful and help understand exactly what is going on.

Problems to be Resolved

The first problem needing resolving is the steam feed. The crack in the boiler found after the final set of experiments were taken shows the results gained in this thesis are invalid and have little meaning in characterising the operation of the CAPE FICFB gasifier. The circulation of bed material and char is a major problem in being able to operate the plant at high temperatures. Further in depth investigation including hot and cold testing of the bed material circulation rates at operating temperature. The use of a transparent stand pipe will make testing this area much easier. To try to increase the gases exposure to higher temperatures and increase the char and bed material mixing the side entry of the bed fuel feed is a high priority. It is however my recommendation that changes are only made one at a time so the relative effect of each can be seen.

5.2 Tar Analysis

The tar samples taken during gasifier operation were analysed by Sean Devenish from the Department of Chemistry at University of Canterbury, who is developing the analysis method. Identification of the compounds contained in the tar material has proven difficult because the mass spectrometer detector in the chemistry department has been out of use. A standard mixture of PAH compounds was purchased and used to identify the main components using HPLC and IR spectroscopy. The unique IR spectra of each compound in the standard, was used with the real tar mixture to identify the compounds. Figure 5-6 is the HPLC print of the compounds washed from the SPE columns with di-chloro methane (DCM), with the identified compounds laboratory on the diagram. Figure 5-7 shows the chromatogram of the compounds washed from the SPE columns with iso-propanol. From these figures there are several major compounds still requiring identification, they are not labelled in these diagrams. Much more work is required in this area including identification of the remaining unidentified compounds and calibration of the method so it can be used quantitatively to give a value for the product gas tar concentration.

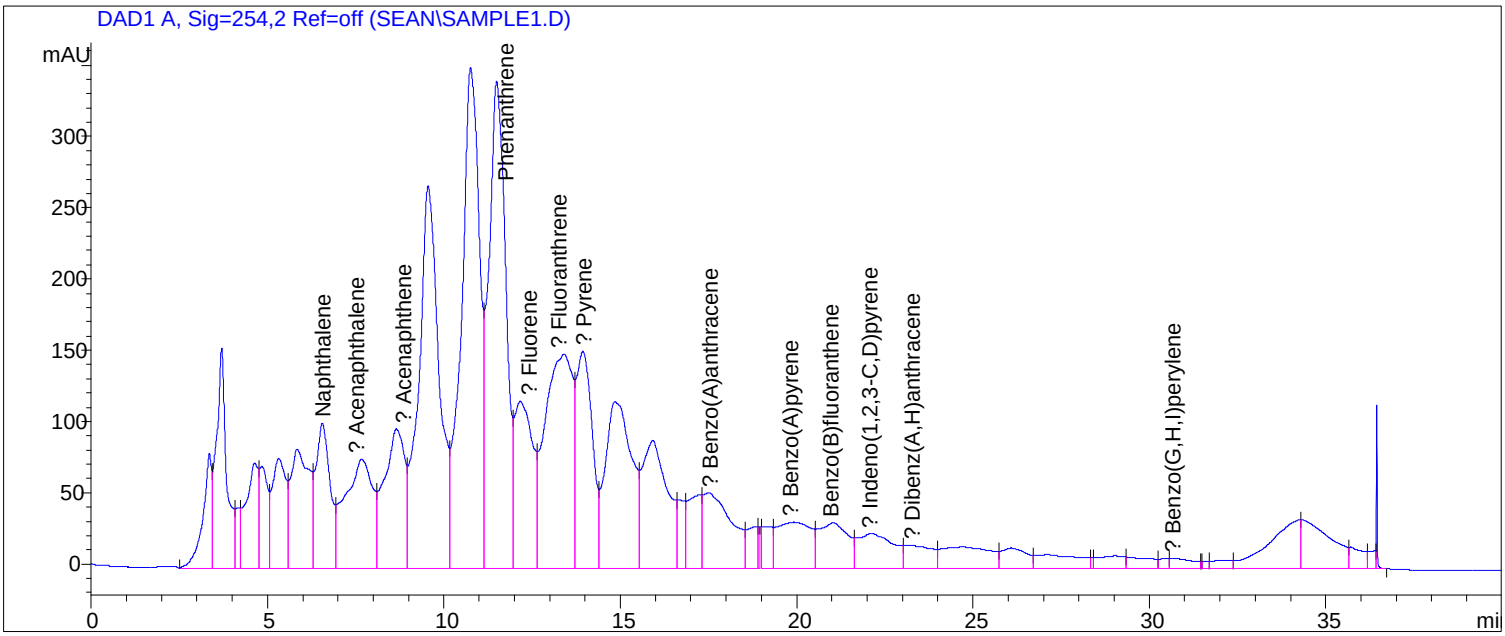

Figure 5-6. HPLC analysis of analyte washed from SPE columns with DCM. Question marks denote compounds identified from the standard PAH mixture.

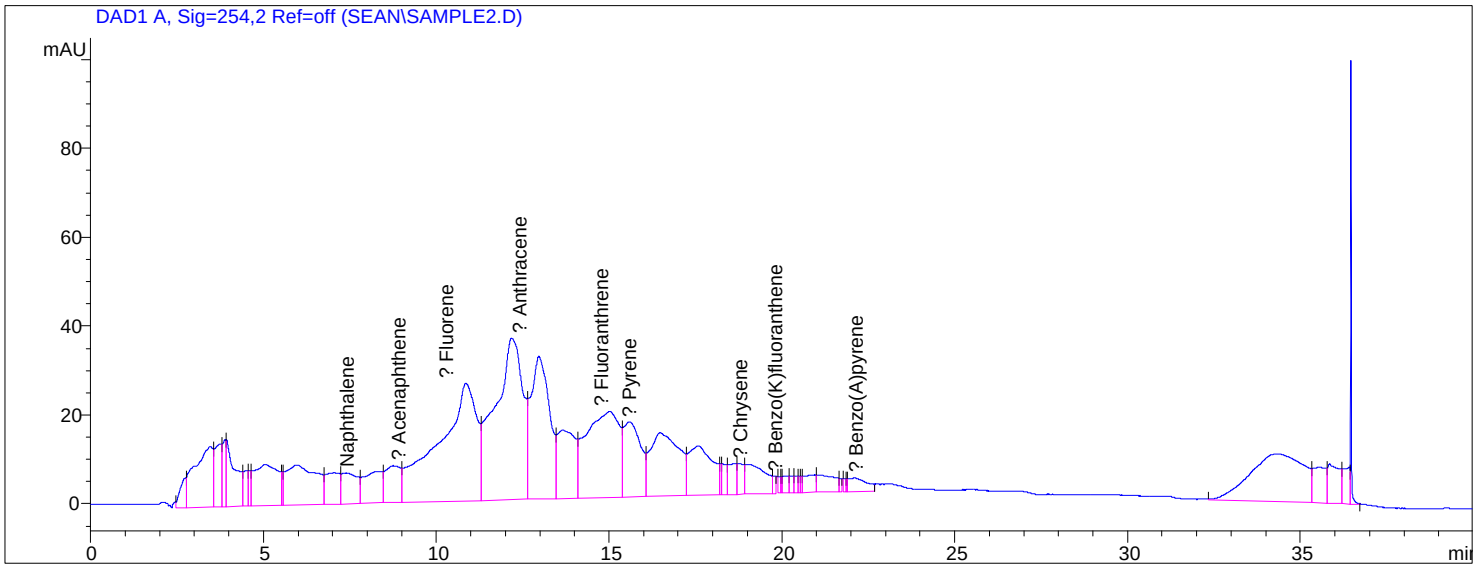

Figure 5-7. HPLC anaysis of analyte washed from SPE columns with IPA solvent. Question marks denote compounds identified from the standard PAH mixture.

6 Page Macrae Updraft Gasifier Results and Discussion

Page McRae is a heavy engineering fabricator situated in Mount Manganui. They employ approximately 100 staff, supplying local and international markets. Design and fabrication for the forestry, oil and gas, dairy and marine industries have been the main stays of the business since it began in 1955 (Macrae 2006). However the current NZ energy situation has led to investment in the development of a counter current (up-draught) moving bed gasification system to fuel boilers to raise process steam.

Figure 6-1. Page Macrae Up-draught Gasification Process. Fuel is fed into the feed hopper at the left hand end. The handling system feeds fuel to the reactor which sits below the flare in the centre. Product gases produced are combusted in the boiler at the top right of the picture.

A 1.7 MW_{th} gasifier fuelling a boiler is being operated at the Carter Holt Harvey (CHH) Laminated Veneer Lumber (LVL or plywood as it is more commonly known) site in Mount Manganui with the aim of being able to give an alternative fuel for boilers currently fired with natural gas (Coulter 2005). The gasifier is fuelled with waste plywood and green wood chip streams from the plant. The aim of the series of experiments is to test the influence of

different operating parameters on the gas composition and complete a mass balance on the plant and determine the efficiency of the gasifier. The data collected during the experiments is also to be used in system modelling.

6.1 Gasifier Process Description

Fuel in the form of the plywood chip, veneer trim (green chip) or debarker waste (mixture of bark and green chip) is fed to a large hopper which holds approximately 24 hrs of feed. A conveyor elevates the fuel into a surge hopper controlled with on/off level sensors controlling the feed. The surge hopper feeds a screw which feeds the fuel on to the top of the bed; the screw speed is controlled with a radar level probe on the gasifier bed height. The gasifier reactor itself is a refractory lined bed approximately 1m in diameter, insulted on the outside with ceramic fibre and aluminium cladding. The fuel bed level is maintained approximately 1.5 m high above a specially designed mechanical grate system in the bottom of the reactor. Ash and un-reacted char fall through the grate into a water bath and are carried out of the reactor.

The oxidant used in the system is air. This is fed at two different points in the system, through the bottom of the bed as primary air and above the bed as secondary over fire air. The primary air is used to control the amount of gas being produced and combusted in the boiler and hence the steam pressure. The gas produced can also be flared to a stack when steam pressure reaches a maximum cut out point. Over fire air is used to combust some of the gas produced to maintain the product gas temperature fed to the boiler and decrease fouling in the feed ducts. The reactor is operated at negative pressure controlled with an induced draught (ID) fan on the exit of the boiler. The product gas produced is combusted in a specially designed afterburner to heat a boiler. The boiler itself is a triple pass configuration controlled with a DURAG photocell flame management system and is certified to be unattended.

6.2 Experimental

Analysis

The methods described in previous sections using the gas cleaning train and micro gas chromatograph were used in this system. A slip stream of the product gases were taken drawn from the duct leaving the top of the reactor going to the afterburner, into a simplified version of the clean up train. The product gases were bubbled through a series of 4 wash bottles filled with iso-propanol solvent in a water bath. Entrained water and solvent were condensed out of the gas in two wash bottles filled with glass wool submerged in a mixture of dry ice and acetone. The cleaned gases were pumped through the system using the Capex V2 diaphragm pump and fed to the micro G.C. at 10 psi to be analysed using the method described in the previous section.

The moisture content of the product gas was found by sucking a known volume of product gas through a cold trap. The same method is used here as in previous experimentation. Five litres of gas is drawn through the cold trap by emptying a cylinder filled with water. The cold trap is filled with glass wool to give a surface for moisture to condense on. The trap is cooled by submerging it in a thermos flask filled with a mixture of acetone and solid carbon dioxide at a temperature of -70°C. By heating the cold trap at 105°C for 20 minutes the moisture was driven off the change in mass is assumed to be the mass of moisture. The cold trap is then washed with solvents to remove the remaining tar compounds from the surface and glass wool. The change in mass after this step is assumed to be the tar fraction.

Fuel samples of plywood chip and veneer trim, along with samples of the waste ash and char, were sent to CRL Energy Ltd for proximate analysis. Ultimate analysis was carried out by Chemsearch Otago. The ash was analysed by Spectrachem Analytical. The gasifier fuel hopper is loaded by the workers from the CHH plant, they keep a log of the fuel added to the gasifier during the week along with a log of the waste ash and char which they also remove. The waste ash and char removed was also measured using a bucket and stopwatch periodically to give an idea of daily variance in the ash and char flow. These were used to calculate the overall carbon conversion.

The program used to run the computer controlling the gasifier does not allow the input and output data to be used by any other program or saved. This made it very hard to be able to look at trends accurately. Screen prints of the short term trends, up to 6 hours long had to be taken to give an indication of what had happened. These gave an indication of the temperature profiles and variations with time and other variable changes.

Gasifier and Operation

For safety reasons the plant can not be operated in a totally manual mode to test the effect of different variables on each other and the gas composition and calorific value. For a reference point, the gasifier was operated under normal conditions to get a set of base line readings for comparison. The set points for controllers were changed and the change in fuel composition and the variation in other variables were monitored. The variables changed were:

- Primary air flow-rate
- Fuel feed rate
- Bed height
- Grate water height

As the variables were altered the above parameters were monitored and plotted to see the effects of each variable on the parameters monitored.

6.3 Results and Discussion

The Page Macrae gasifier has been in development for six years in the hope of developing a piece of equipment that can operate unsupervised. In July 2006 development reached the stage where the plant could be operated continuously during the week with only operator intervention during the day. For two weeks up to the 4th of August 2006 the following sets of results were collected operating the up-draught gasifier on plywood chip for which the composition is shown in Table 6-1.

The fuel samples were analysed by CRL Energy Ltd and the results of this analysis are shown in Table 6-1. Analysis of the ash was carried by Spectrachem Analytical. They were unable to determine 90% of the components of the ash for which the analysis is shown in Table 6-2. Despite not being able to identify most of the sample observations of the ash melting at 815°C with a slightly blue colour which suggested the presence of borate most probably combined with sodium. The report issued by CRL Energy Ltd can be found in Appendix A. The phenol

formaldehyde resin, the flour used as a resin extender holding the veneers making up the plywood board together or the preservatives used on the veneers are most likely the source of these un-identifiable compounds. Borax is a common preservative that contains $Na_2B_4O_7$. Bifenthrin ((2-methyl-1,1-biphenyl-3-y1)-methyl-3-(2-chloro-3,3,3-trifluoro-1-propenyl)-2,2-dimethyl-cyclopropanecarboxylate) is an insecticide additive used. It is most likely that a mixture of the resins and flour used as an extender and the residue produced from these is at least partially responsible for the grate blockage problems explored later in this section.

Table 6-1. Proximate and ultimate analysis of the wood samples from Page Macrae fuel feed and ash waste an oven dried basis in wt%. The analysis was carried out by CRL Energy Ltd.

	Sample #1, Green Chip	Sample #2, LVL Chip	Sample #3, Ash and Char Waste
Moisture (Wet Basis)	45.6	30.3	71.8
Proximate Analysis			
Ash	0.3	1.2	31.3
Volitile	83.3	77.7	13.2
Fixed Carbon	16.4	21.1	55.5
Ultimate Analysis			
Carbon	50.3	50.8	65.8
Hydrogen	6.12	6.22	0.8
Nitrogen	<0.2	<0.3	<0.3
Sulphur	<0.01	<0.01	0.03
Oxygen (by difference)	43.3	41.8	2.1
Gross Calorific value (MJ/kg)	20.2	20.36	22.09

Table 6-2. X-ray Fluorescence major oxide analysis of ash sample by Spectrachem Analytical. Results expressed in weight % at 815°C.

SiO_2	7.11
Al_2O_3	2.17
Fe_2O_3	1.03
CaO	6.71
MgO	2.30
SO_3	0.60
K_2O	6.93
Na_2O	46.04
MnO	0.33
TiO_2	0.11
P_2O_5	1.27
Total	74.60
Loss on Ignition*	3.83

* Loss on ignition on sample as received, at 815°C for 1 hour.

The fuel added to the gasifier during the period was recorded along with the waste ash material removed by the grate extraction system. The grate has a tendency to become blocked and then release large amounts of ash and char all at once making it hard to record the amount of char being produced under different conditions to find the carbon conversion. However by measuring the fuel fed and ash recovered over the period of testing an approximate carbon conversion was found. During the testing 141.5 m^3 or 37 000 kg of LVL chip fuel was fed to the gasifier and 3.6 m^3 or 890 kg of ash waste was recovered. The compositions of both of these are shown in Table 6-1. These figures give an overall carbon conversion for the testing period of 98 % by weight.

In the gasifier wood fuel is partially oxidised in the presence of air. It follows that the flow-rate, composition and moisture content of the gas produced should be influenced by changes in the air flow-rate. Figure 6-2 shows the gas composition plotted along side the primary air mass flow-rate both against time and Figure 6-2 shows the composition against time for a step change in the primary air mass flow-rate. Figure 6-2 shows the gasifier operating under normal automatic control with the air rate being used to control the boiler steam pressure. As the primary air is reduced the amount of gas produced is also reduced so the heat in the boiler is also reduced. When the boiler pressure gets too high the product gases are vented to stack, this normally only occurs when the steam demand of the LVL plant is very low. When the product gases are flared to stack, the air flow-rate reduces to 1.5 kg/min to maintain the temperature in the bed while not wasting fuel.

The nitrogen content of the product gas is quite clearly linked to the air flow-rate in Figure 6-2. A reduction in the primary air flow-rate causes an instantaneous reduction in the nitrogen composition; however this does not seem to be permanent. The air mass flow-rate reduces from 7 kg/s to 6 kg/s between 13 and 14 hours after mid-night, after the initial drop in the nitrogen content it increases back to its initial concentration. Similar effects can be seen in the concentrations of hydrogen, carbon dioxide and methane which initially increases then recovers and carbon monoxide which decreases then recovers. To get a continual consistent gas flow and composition, changes to the operation of the LVL plant are required including smoothing of the heat load. Alteration of the operational procedures on the plant would help smooth out the gasifier operation but recommendations to this are beyond the scope of this thesis and will require further work.

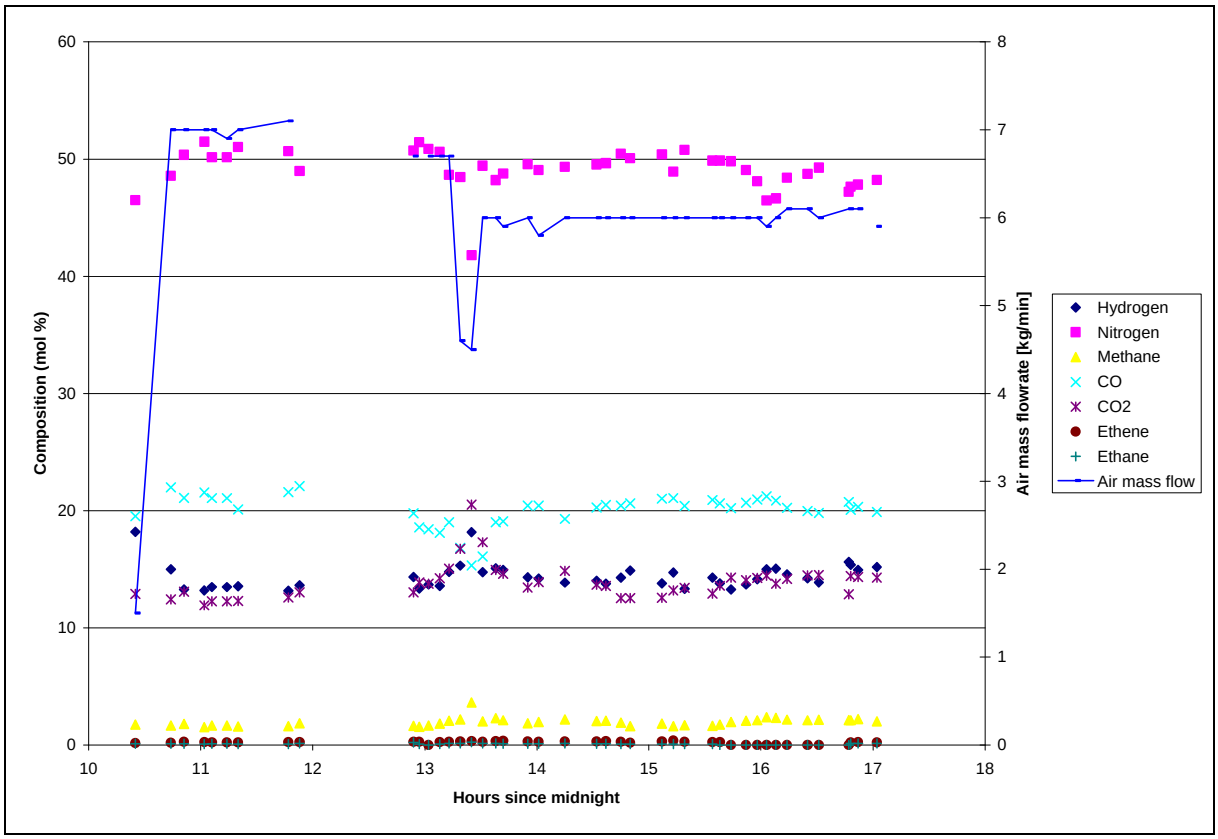

Figure 6-2. Page McRae gasifier composition and air flow-rate to the gasifier show a weak relationship between the two. Data recorded on 3/8/06.

Figure 6-3 supports the idea that the composition is not dependant on the primary air flow-rate; however, there are other mechanisms at work causing deviations from the baseline composition. Thus Figure 6-3 is not a good representation because the reduction in primary air flow-rate to the gasifier from 7.5 kg/s to 4.5 kg/s caused the grate to block up and the water level in the grate had to be increased. This may have had some unexpected effect on the composition of the product gas. As we can see in Figure 6-3 when the water level is increased 15.5 hours after mid-night there appears to be no pattern in the changes to the composition, other than a drop in the carbon monoxide concentration from 20 to 16 mol.%. It is most likely that the grate had blocked perhaps at around 15 hour since mid-night when the composition changes unpredictably, before the water level increased. This caused unnoticed changes in the control variables which led to the changes in composition. From the logs there is no obvious drop in the product gas temperature as it fluctuates ±50°C under normal operation due to changes in the air feed rate.

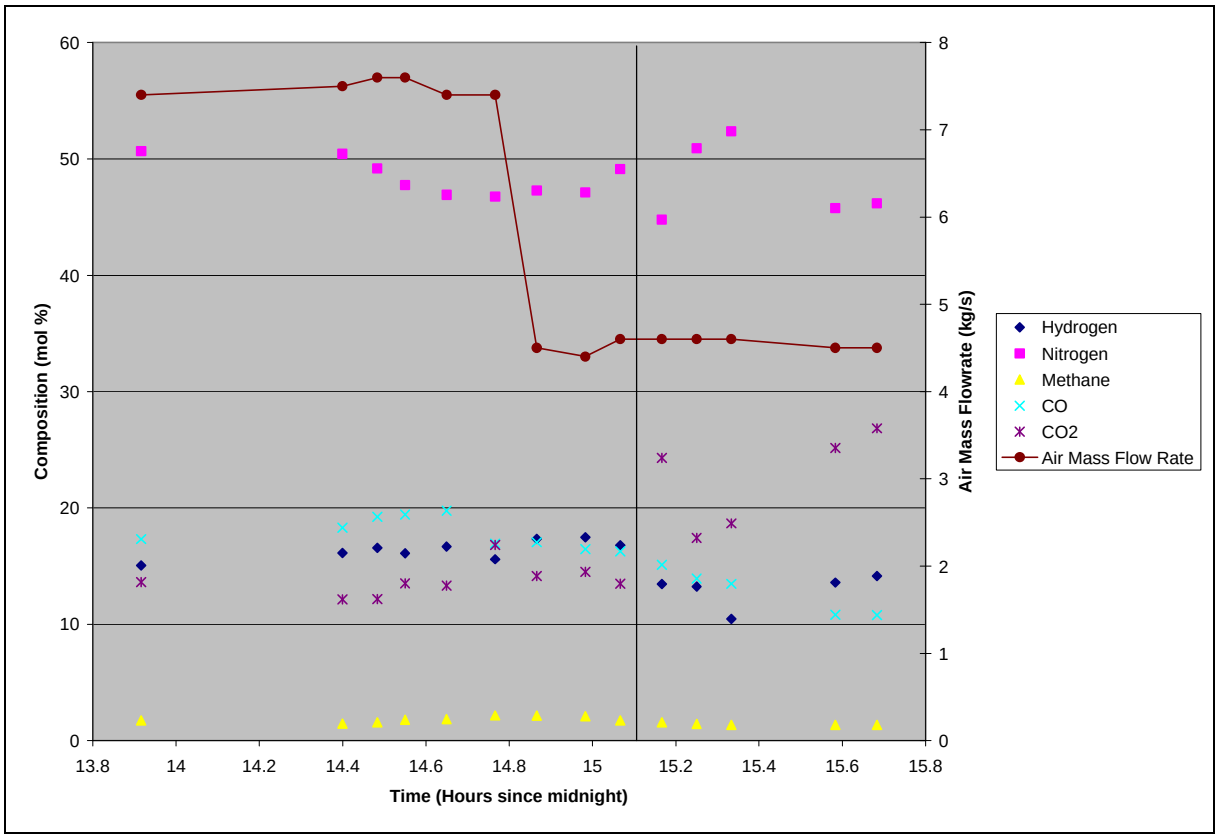

Figure 6-3. Composition and air mass flow-rate again time for a step change in the air flow-rate. Data recorded on the 4/8/06. Line shows when water level below grate was increased.

The bed height is not related to the composition of the gas produced for bed height set points of 1000 mm, 1300, and 1700 mm as shown by Figure 6-4. Figure 6-2 showed that the changes seen in the composition are more likely related to changes in the primary air flow-rate. The constant composition indicates that the reactor is in equilibrium.

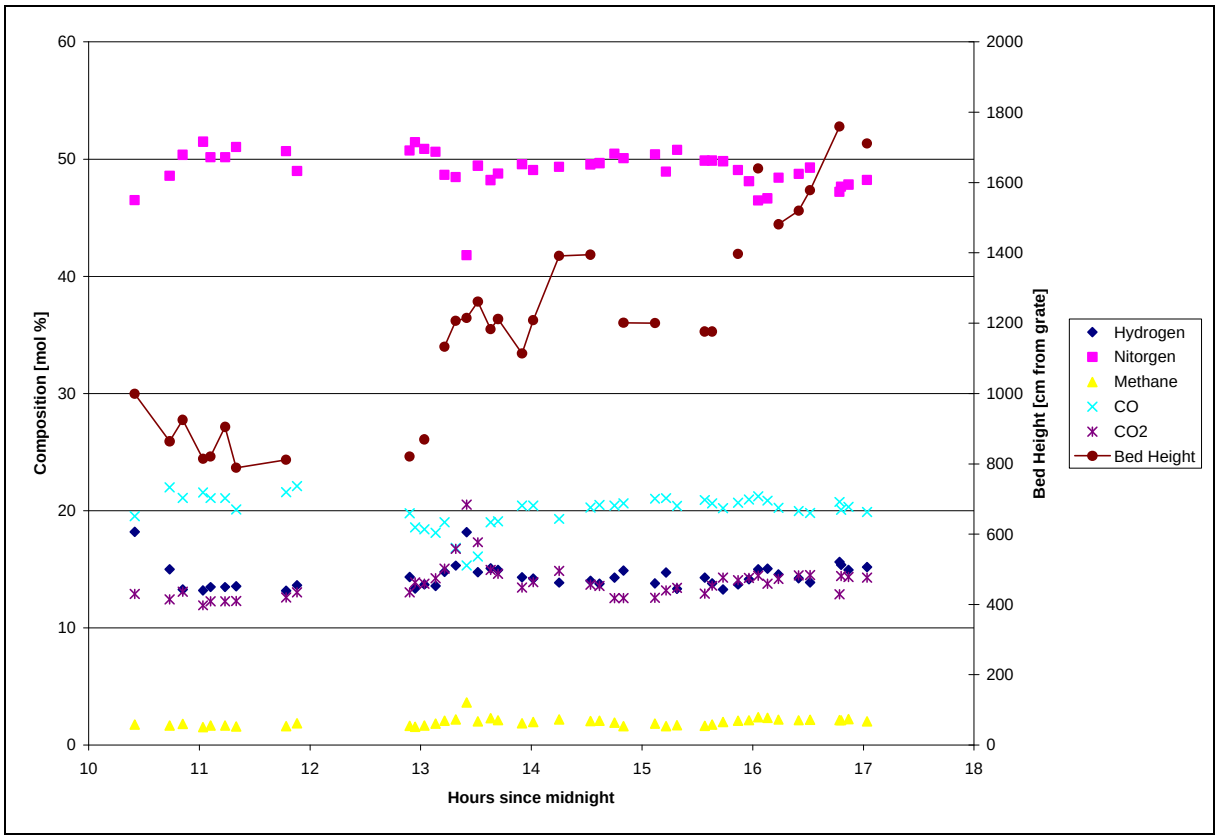

Figure 6-4. Page McRae gasifier composition against bed height shows little variation in composition with changes in bed height. Data recorded on 3/10/06.

The moisture content of the product gas does not have a noticeable effect on the product gas composition as shown in Figure 6-5. The moisture content varies between 30 and 49 mol.% (wet basis); however, the dry gas composition shows no significant change that can not be explained by other factors. The problem with drawing a link between the moisture content and the gas composition is the sampling method for moisture content meant the samples took 5 minutes to take and only one sample port meant that gas composition could not be taken at the same time. The moisture content is directly related to the amount of fuel being added as shown by Figure 6-6 and Table 6-3. As the fuel feed rate increases the moisture content of the product gas increases. This can be explained by the extra moisture added in the fuel being dried out of the fuel quickly as it enters the hot bed. Figure 6-6 shows a log of the bed height, fuel screw speed and product gas temperature against time. Between 13:00 and 15:30 hours since mid night the set point of the bed was set at 1300 mm from the grate however there was about 300 mm variation in the height. The 300 mm variation corresponds to 0.8 m³ of fuel or 170 kg, the moisture content of the plywood chip fuel, on a wet basis, is 30.3 % which corresponds to 50 kg of water added to the system. With the 300 mm variation in height it is assumed no fuel is used up in the mean time, so the reality is more water than this is added.

The log also shows the product gas temperature is closely related to changes in bed height and hence moisture content of the product gas, due to the heat requirements of the phase change drying the fuel. The gas temperature must be maintained above 550°C to prevent tar formed during gasification condensing on surfaces (Neeft, Knoef et al. 2002). Figure 6-6 shows the temperature of the gas gets as low as 350°C and is less than 550°C a considerable proportion of the time. Tar condensation in the duct leaving the top of the gasifier leads to restriction of the gas flow and eventually total blockage. Controlling the bed level so large amounts of fuel are not dumped into the reactor reducing the temperature should in turn reduce tar condensation. Figure 6-6 shows the feeder screw is slow to ramp up in response to changes in the bed level. Tighter control of the bed height would lead to less fluctuation in the moisture content and hence the temperature of the product gases. This may in turn reduce the build up on the walls of the product gas duct and stop it blocking.

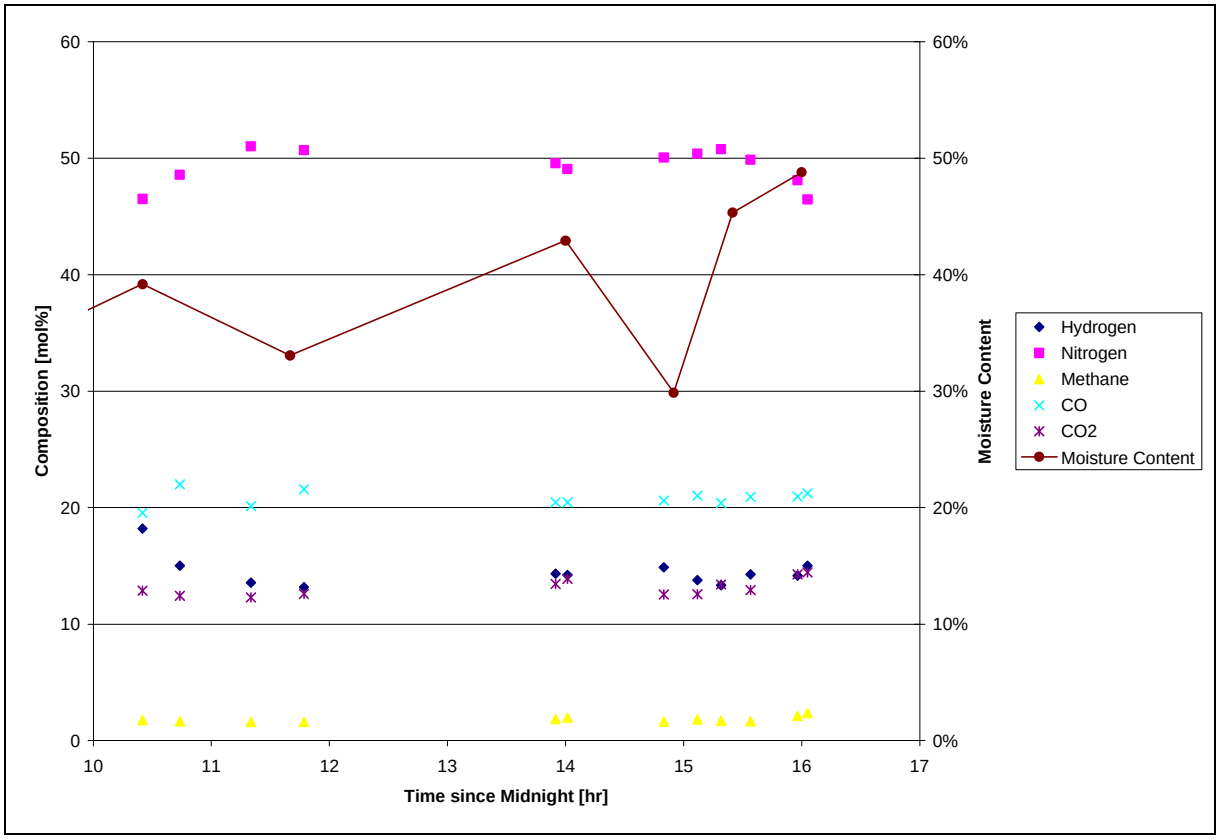

Figure 6-5. Moisture content of product gas and dry gas composition plotted against time for the Page Macrae up-draught gasifier. Data recorded 3/10/06.

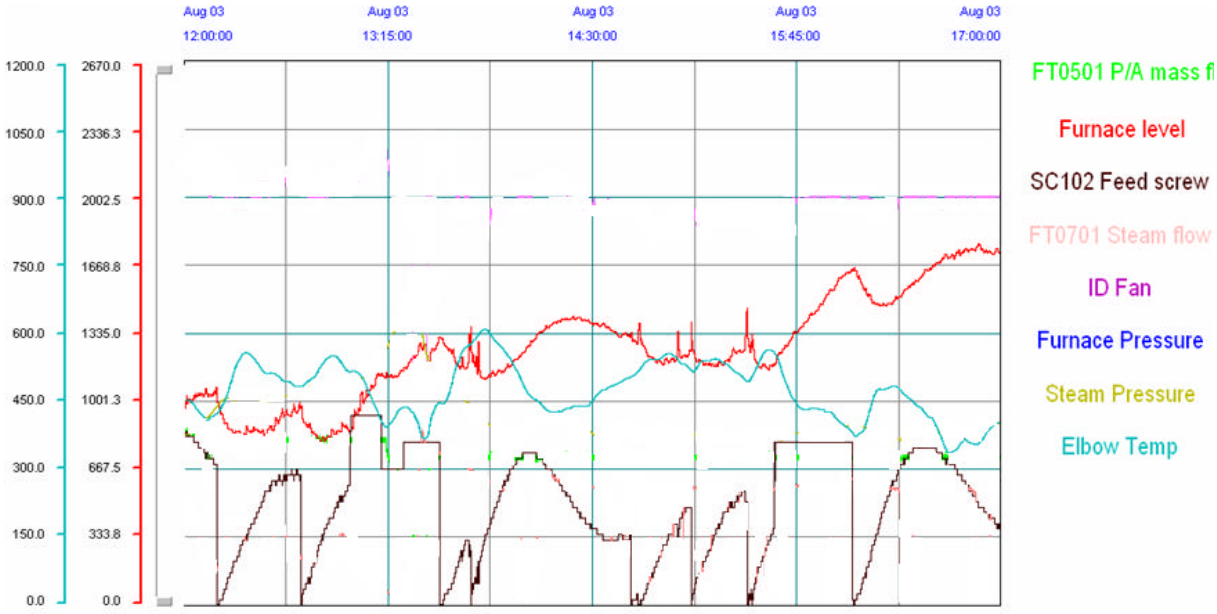

Figure 6-6. Log of the bed height (furnace level, cm from the grate), feed rate (SC Feed screw), and product gas temperature (Elbow Temp, in °C) which shows their relative effect on each other.

Table 6-3. Moisture content of the product gas compared with change in bed height.

Time	10:25:00	11:40:00	14:00:00	14:55:00	15:25:00	16:00:00
Hours since mid-night	10.4	11.7	14.0	14.9	15.4	16.0
Moisture (mol%)	39%	33%	43%	30%	45%	49%
Bed height change	increasing	decreasing	increasing	steady	beginning to increase	Big increase

Objective four of the BIGAS project at the University of Canterbury includes Gibbs energy equilibrium modelling of the gasifier composition. The model developed by Jack Rutherford has been used to describe the Page Macrae gasifier, Table 6-4 shows the gas compositions predicted by the equilibrium modelling of Rutherford compared with the averaged results for the Page Macrae up-draught gasifier. A description of the modelling approach can be found in the thesis submitted by Jack Rutherford for his Masters in Chemical and Process Engineering (Rutherford 2006).

Table 6-4. Comparison of Equilibrium modelling results with the range of measured gas compositions for the Page Macrae up-draught gasifier (Rutherford 2006).

	Measured	Equilibrium			
Bed Temperature (°C)	525 to 625	570	620	650	Modified (650°C)
Dry Composition					
Hydrogen (mol %)	13 to 19	21	20	19	19
Methane (mol %)	2	3	1	1	1
Carbon Monoxide (mol %)	13 to 19	21	20	19	21
Carbon Dioxide (mol %)	12 to 14	22	17	13	13
Nitrogen (mol %)	45 to 51	45	46	46	46
Lower Heating Value (MJ/Nm³)	4.1 to 5.1	4.1	4.4	4.6	4.6
Efficiency (LHV)	34 - 52%				
Water Content (mol % wet basis)	30 to 50	20	13	8	30

The Page Macrae gasification system operates close to equilibrium as indicated by the results in Table 6-4 above and Figure 6-4 which shows consistent gas composition with changes in bed height from 900 mm to 1700 mm. The moisture content of the product gas is not predicted accurately by the equilibrium model. Although the temperature that predicts the gas composition best is 650°C and is 25-90°C hotter than the recorded bed temperature, the recorded temperature may not be a true representation of the reaction temperature due the temperature gradient in the reactor. It also indicates that the composition of the final gas is determined by hotter regions of the combustion zone of the gasifier. The moisture content of the gas is not predicted accurately by the model, but in a modified model where it is assumed only dry fuel reacts and moisture is dried off separately the moisture is much more accurately modelled, as can be seen in the last column of Table 6-4. This suggests the most likely explanation for this is the fuel bound moisture is dried when the fuel drops from the feeder on top of the bed in the low temperature drying zone having little reaction with the gas composition, since it does not pass through the high temperature char zone. The efficiency of the gasifier was 34 – 52% (LHV) and was very dependant on the moisture content of the product gas. At high product gas moisture contents, up to 50 mol% the plant efficiency was at a minimum. The large amount of carbon leaving the gasifier in the ash also reduces the plant efficiency. Tars are not taken into account in the conversion efficiency calculation but are mostly combusted in the boiler so the actual efficiency is higher than that calculated. It is not uncommon for 10-20% of the input energy of the wood to be converted to tar material in counter current moving bed configuration (Milne, Evans et al. 1998).

7 Conclusions and Recommendations

7.1 CAPE FICFB Gasifier Operation

The CAPE (Chemical and Process Engineering) gasifier produces a product gas with lower heating values of 10400-12100 kJ/Nm³ under the current operational constraints. There is room to improve and optimise this value by implementing the proposed modifications discussed in the previous section. The calorific value of the gas produced by the CAPE gasifier is lower than that produced by the Vienna gasifier as a result of the higher nitrogen content in the product gases. Under current operation air is fed to the siphon and chute to fluidise them, increasing the amount of nitrogen in the gases. By fluidising the siphon and chute with steam instead of air, the calorific value can be improved. The methane, ethane and ethene concentrations of the product gas are higher than the Vienna results suggested. This is attributed to the lower operating temperatures of the CAPE gasifier and differences in the feed system. At the CAPE plant, fuel is fed directly on top of the bubbling fluidised bed in the gasification column as opposed to through the side of the bed at the Vienna plant. It is suggested that by feeding fuel directly on top of the bed, the gases formed are not subjected to the temperature of the bed material for a sustained period, which means that the amount of thermal cracking is reduced, hence explaining the higher concentrations of hydrocarbons. The temperature of the gasification column is limited by the heat transfer rate from the combustion column. The heat transfer rate is determined by the bed material circulation rate and the temperature of the bed material being circulated. Under current operation, the bed material circulation rate is restricted owing to the cyclone and siphon system not separating bed material from the combustion gases at the same rate as it is fed; hence, bed material is often blown through the heat exchangers. It is suspected that interactions between char and bed material are causing the problem, but further investigations are required. Not enough data has been collected to draw conclusions on the effects of different manipulated variables on the product gas composition. The steam to fuel ratios need to be reassessed when the boiler leak has been repaired or a new steam source is in place.

7.2 CAPE FICFB Gasifier Recommendations

The operation of the CAPE gasifier has highlighted several issues that require attention for long term operation. To reduce the tars and hydrocarbons, thermal cracking mechanisms are one possibility and increasing the temperature to which product gases and fuel are exposed is one option. By feeding bed material through the side of the bed, the fuel will be exposed to higher temperatures for longer periods of time. Increasing the temperature of the bubbling bed in the gasification column will also help. The bed material circulation rate is the problem. To increase this rate, investigation into bed material particulate size and cyclone and siphon design are required. The cause of the bed material segregation in the bed cyclone and siphon could hold the key to increasing the circulation rate. During operation another problem with the system has been the steam supply. Not enough steam could be fed to the siphon and chute regions to fluidise them during operation; it is therefore recommended that a new steam supply is investigated to replace the leaking boilers. Tar analysis is a vital part of the analysis and comparison of the gasifier reactors; however, the development of the tar analysis process and compound identification has held back this project and it requires further urgent attention. Finally, the product gas moisture content is hard to measure and obtaining samples takes a long period of time. Another measurement method or at least new port is required to remedy this problem. Flow-rate measurement of the product gas flow is required to evaluate the variation in the amount of gas produced over time.

7.3 Page Macrae Gasifier Operation

The overall carbon conversion of the plant during the testing period was measured to be 98%; this means the gasifier is producing a waste stream containing useful carbon that requires disposal and reduces plant efficiency. The lower heating value of the gas produced by the system is 2700-5200 kJ/Nm³ and averages 4700 kJ/Nm³ on a time weighted basis. This is too low to use in a gas engine or turbine for electricity generation but suitable for its current application in a steam boiler. The calorific value is lowest when the primary air rate is low and a large proportion of the gas produced is combusted in the reactor by the over fire air supply, which is constantly supplied at the same rate. The gasifier operates close to equilibrium predicted by the Gibbs energy model developed by Jack Rutherford however it does not predict the moisture content of the product accurately. The moisture content of the actual gas varies from 30-47 vol%, depending on the amount of fuel entering the reactor,

compared with 20-8 vol% predicted by the model. If fuel moisture is assumed to dry on top of the bed, and not react with the fuel and oxidant, the modelled composition matches much more accurately. When large amounts of fuel are fed into the reactor, the moisture content is high because of the drying. This in turn drops the product gas temperature and increases the risk of tar condensation in the duct work. Better control of the bed height by adjusting the controller parameters would decrease the amount of fuel being dumped into the reactor at once by reducing the ramping effects of the integral control.

7.4 Gasifier Comparison

There are many comparisons that can be drawn from the analysis of both the Page Macrae up-draught and CAPE FICFB gasifiers; however, it is important to realise that the two processes have been designed for different applications. The Page Macrae gasifier has been designed for steam generation and low capital cost to supply gas fired steam boilers and hence has a very simple design since gas quality is not such a concern. The CAPE FICFB gasifier is designed with electricity and high value products in mind and so has a more complex design to produce a high quality product gas. The CAPE gasifier produces a gas with a lower heating value (LHV) of 10400-12500 kJ/Nm³ compared with the 4100-5100 kJ/Nm³ produced by the Page Macrae up-draught gasification system. The heating value of the CAPE gasifier makes it much more suited to gas turbine and engine applications. However, the simple design and hence low investment required for the Page Macrae process makes it economically competitive for steam generation whilst also exceeding emissions restrictions.

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A. Appendix A

Fuel Analysis

This section outlines the fuel analysis carried out by CRL Energy Ltd on an as received basis. The proximate and ultimate anlysis of the pellet and chip fuels used in the CAPE FICFB gasifier in Figure A-1. Report on proximate and ultimate analysis of the pellet and chip fuels used in the CAPE FICFB gasifier; carried out by CRL energy on as recieved basis.The following is a letter written by Ian Gilmour to Harraway and sons of Dunedin. This contains the ash analysis of the fuels used in the gasifier. The analysis of the Page Macrae plywood chip (chipped waste board), veneer trim (chipped green wood waste) and ash was undertaken by CRL Energy Ltd and the results are shown in Figure A-2. The Analysis of the ash composition was organised by CRL Energy Ltd but under taken by Spectrachem, the report on this work is shown in Figure A-3 and the table on the following page.

Pellet Analysis

CRL Energy Ltd

INTERIM REPORT OF ANALYSIS

Date Received: 26-Aug-05

Client: Canterbury University

Description: Wood Chip pellets and Husk samples supplied by client.

CRL Energy Ltd Reference:			76/050	76/051	76/052
Customer Reference:			Sample#1 Chips	Sample#2 Pellets	Sample#3 Husks
Analysis - As Received Basis					
Moisture	ISO 5068	%	52.6	8.0	9.9
Ash	ASTM D1102	%	0.2	0.4	2.6
Volatile	ISO 562	%	39.8	77.4	73.8
Fixed Carbon	By Difference	%	7.4	14.2	13.7
Gross Calorific Value	ISO 1928	MJ/kg	9.53	18.63	17.08
Carbon	micro analytical	%	24.3	47.2	43.7
Hydrogen	micro analytical	%	2.87	5.35	5.07
Nitrogen	micro analytical	%	<0.1	<0.2	0.56
Sulphur	ASTM D4239	%	0.01	0.01	0.06
Oxygen	By Difference	%	20.0	38.7	38.1
CHN determined by Chemsearch Otago University					
Analysis - Dry Basis					
Ash	ASTM D 1102	%	0.4	0.4	2.9
Volatile	ISO 562	%	84.0	84.1	81.9
Fixed Carbon	By Difference	%	15.6	15.4	15.2
Gross Calorific Value	ISO 1928	MJ/kg	20.10	20.25	18.95
Carbon	micro analytical	%	51.2	51.3	48.5
Hydrogen	micro analytical	%	6.10	5.81	5.63
Nitrogen	micro analytical	%	<0.2	<0.2	0.62
Sulphur	ASTM D4239	%	0.02	0.01	0.07
Oxygen	By Difference	%	42.3	42.4	42.9

Signature:

Date of Issue: **13-Oct-05**

Grant Murray

Laboratory Supervisor

THIS REPORT MUST NOT BE QUOTED EXCEPT IN FULL

Distribution:

Dept of Chemical and Process Engineering, PB 4800, CHCH ATTN: Ian Gilmour

CRL Energy Ltd, Laboratory

Figure A-1. Report on proximate and ultimate analysis of the pellet and chip fuels used in the CAPE FICFB gasifier; carried out by CRL energy on as recieved basis.

Pellet Ash Analysis

		Pellets	Husk
Ash	%wt	0.4	2.9%
Nitrogen;	%wt	< 0.2	0.5% by wt
Sulphur;	%wt	0.01	0.06%

Ash constituents;

		Pellets	Husk
SiO2	% wt	20.50	71.72
Al2O3	% wt	4.66	0.11
Fe2O3	%wt	2.72	0.28
CaO	%wt	24.37	4.37
K2O	%wt	21.51	11.00
Na2O	%wt	1.75	1.04

Comment: The only higher constituent is SiO2, silica, which is a benign compound of soils. This is the most abundant compound on earth so poses no threat.

Trace element analysis;

Comment: There is no comparison with wood pellets here but considering the absolute values these are all in trace quantities of micrograms per gram of material. None of these elements are volatile so they will all be contained in the ash.

Any toxic elements such as Arsenic, Cadmium, Nickel, Lead, Tin, Uranium, Mercury are all at or below the lowest detectable level of measurement. These pose no threat whatsoever to the environment.

Most dominant elements are Iron, Manganese, Copper, Strontium and Zinc still at trace levels but these are all essential elements for life. These are so low as to be considered benign.

CRL Energy Ltd

REPORT OF ANALYSIS

Customer: University of Canterbury

Description: Samples supplied by client

Date Received: 10-Sep-04

Customer Reference:			**Sample # 1 Wood Chips**	**Sample # 2 LVL Chip**	**Sample # 3 Char + Ash**	**Sample # 4 Large Lumps from Char + Ash**
CRL Energy Ltd Reference:			**79/262**	**79/263**	**79/264**	**79/265**
Total Moisture* (As Received)	(Loss on drying at 105ºC)	%	45.6	30.3	71.8	29.6
* On the wet weight of wood basis						
Analysis - Oven Dry Basis						
Ash	(ASTM D 1102)	%	0.3	1.2	31.3	83.4
Volatile	(ISO 562)	%	83.3	77.7	13.2	15.3
Fixed Carbon	(by difference)	%	16.4	21.1	55.5	1.3
Carbon	(micro elemental analysis)	%	50.3	50.8	65.8	17.0
Hydrogen	(micro elemental analysis)	%	6.12	6.22	0.80	0.41
Nitrogen	(micro elemental analysis)	%	<0.2	<0.3	<0.3	<0.2
Sulphur	(ASTM D4239)	%	<0.01	<0.01	0.03	<0.01
Oxygen	(by difference)	%	43.3	41.8	2.1	
Gross Calorific Value	(ISO 1928)	MJ/kg	20.20	20.36	22.09	3.16
C,H,N were determined by Chemsearch Otago						

Date of Issue: 22-Sep-06

Signature:
Grant murray
Laboratory Manager

THIS REPORT MUST NOT BE QUOTED EXCEPT IN FULL

Distribution:
University of Canterbury, Dept of Chemical & Process Engineering ATTN: Ian Gilmour

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SPECTRACHEM ANALYTICAL

CLIENT	:	**CRL ENERGY LTD**		
ADDRESS	:	P O BOX 31-244, LOWER HUTT		
EMAIL	:	g.murray@crl.co.nz	***PHONE***	: 570-3700
FAX	:	570-3701		
ATTENTION	:	GRANT MURRAY	**JOB REFERENCE : SA10271-F**	

CLIENT REFERENCE	:	79/264 Char + Ash
SAMPLE TYPE[S]	:	WOOD ASH
DATE OF SAMPLE RECEIPT	:	19/09/2006 **CONDITION :** POWDERS
ANALYSES CARRIED OUT	:	XRF MAJOR OXIDES ; LOSS ON IGNITION
REPORTING BASIS	:	ASHES - WEIGHT % AT 815°C

The analytical results presented in this report apply to the sample(s) received by SpectraChem Analytical.

Analysis	***Method used***	***LLD***	***Unit***
LOI	*Loss on ignition at 815°C*	0.01	%
Major oxides	*Borate fusion / X-ray spectrometry*	0.01	%

Comments : *These results should be considered qualitative, as we are unable to determine 90-100% of the sample. From the observation that the sample melted with a slightly blue colour at 815°C, and the information it is a wood ash sample we believe the remainder may be borate, most probably combined with the sodium.*
Borax, $Na_2B_4O_7$ is sometimes used as a timber preservative.

XRF preparation and analyses were carried out in accordance with ASTM standard procedure D4326-01.

Date : 21/09/2006 ***Signed :*** Craig Fraser IANZ Signatory

SpectraChem Analytical Limited : 36 Seaview Rd : Lower Hutt
P O Box 38-680 Wellington Mail Centre : Tel. 04 589-6333 : Fax. 04 569-6605 : Email. spectra@spectrachem.co.nz

COAL RESEARCH LIMITED

JOB REFERENCE : SA10271-F

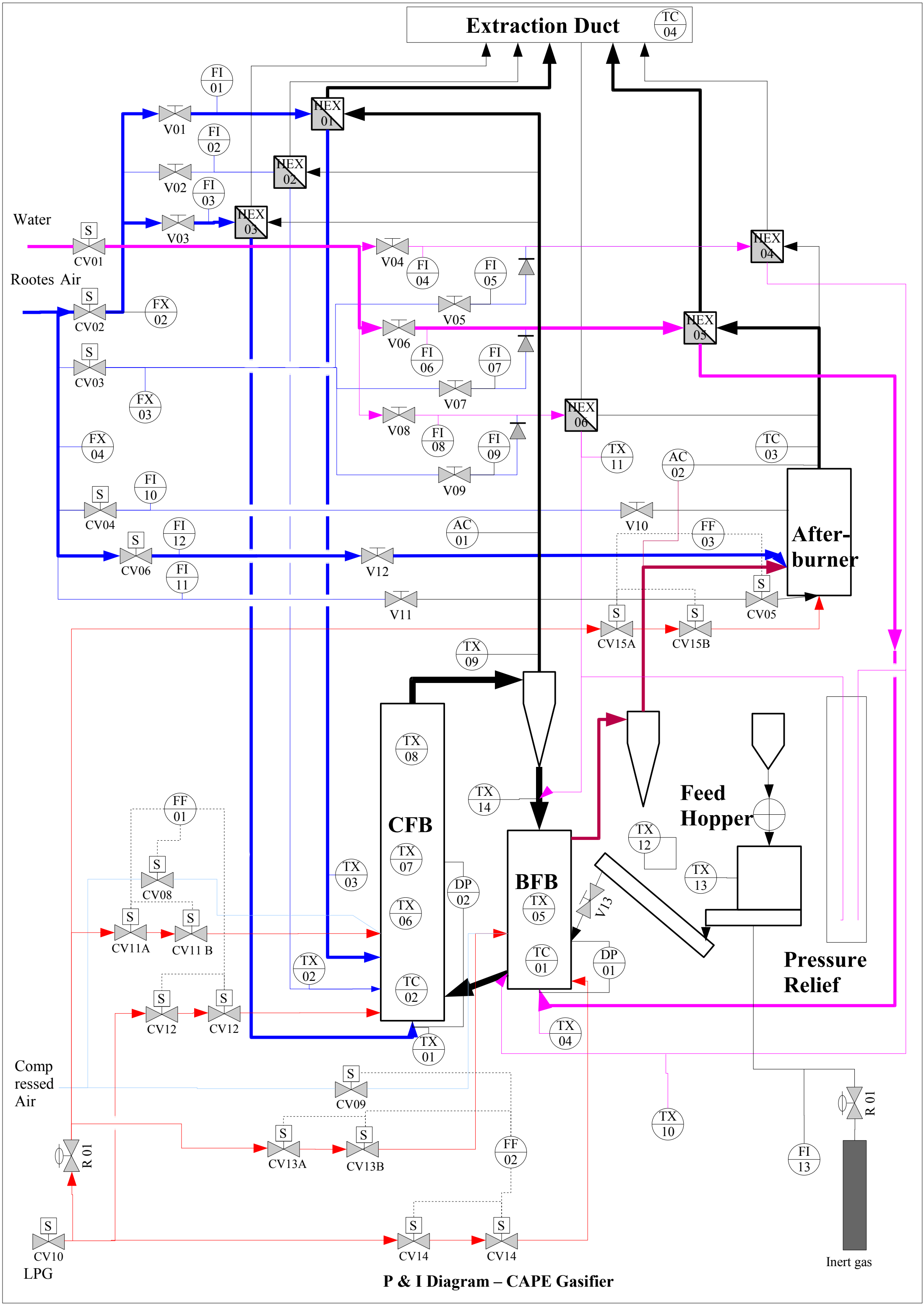
X-RAY FLUORESCENCE MAJOR OXIDE ANALYSES

SAMPLE >	79/264 Char + Ash
SiO2	7.11
Al2O3	2.17
Fe2O3	1.03
CaO	6.71
MgO	2.30
SO3	0.60
K2O	6.93
Na2O	46.04
MnO	0.33
TiO2	0.11
P2O5	1.27
SUM	74.60
LOI	3.83

Results expressed as weight % at 815°C.
LOI = loss on ignition on sample as received, at 815°C for 1 hour.

B. Appendix B

Gasifier Plant Process and Instrumentation Diagram



P & I Diagram – CAPE Gasifier

C. Appendix C

Gasifier Operations Guide

Operational Guide to the CAPE Fast Internal Circulating Fluidised Bed (FICFB) Gasifier

Written by Jock Brown with input from Rick Dobbs, Ian Gilmour and Jack Rutherford

Last Updated: December 2006

Table of Contents

Suppliers

Air line and Fittings: SMC
Controllers: Intech
Kaowool Products: Foreman Insulation
LPG fittings and pipe work: Aquatech
Micro GC Calibration gas: BOC
Nitrogen Purge Gas: BOC
Pipe Fittings and Valves: Taylors or Steel and Tube
Pressure sensors (Bed differential sensors): RS
RTV (Use High Temperature red coloured): Blackwood Paykels or Homershoms (Blackwoods stuff seems to be better).
Sample gas pump: 4ways pumps
Solid Phase Extraction columns: Biolab, NZ
Stainless Steel Parts: Steel and Tube Stainless used to be NZF Stainless
Swaglok fittings and tube: Fluid Systems
Thermocouple Supplies: Homershoms
Thimble Filters (clean up train): Scientific Supplies

Maintenance

Every Run

- Clean up train
- Empty hoppers
- Clean siphon by removing bung in the bottom and checking for wet and agglomerated bed material.
- O2 sensors clear and operational
- Calibrate GC
- Inspect burner flame rods and igniters and check operational.

Monthly

- Inspect totally burner assembly for build up of agglomerated bed material in cones, tars and particulate build up in the inner tubes.

Yearly

- H2 and CO sensor checks
- ? P sensors
- Check all steel work and cyclones; remove and dismantle sections and check all refractory for cracks. Photograph and compare with previous photographs.
- Test burner flame failure systems and have inspected (follow standard for LPG systems)

Dismantling and Assembly of Refractory Sections

Always remove the distributor and bottom section of the combustion column (circulating fluidised bed) first. It is best if you have two people for this as it is a slow process lowering the sections on the block and tackle with only one person. Before removing the bottom distributor clean the floor underneath the columns so when the distributors are removed so bed material is not contaminated with rubbish on floor.

Equipment and supplies required for disassembling refractory sections of gasifier:

- Block and Tackle
- D Shackles
- Kaowool blanket
- Kaowool rope
- RTV

Procedure for removing sections:

- Loosen and remove all but two bolts on opposite sides of column as a safety measure.
- Work the RTV free with a wedge and sharp edge so the column section rests on the bolts.
- Attach D shackles to two opposing holes
- Attach block and tackle to grate in mezzanine floor above and to D shackles.
- Take weight of section with block and tackle and remove bolts
- Lower section to the floor lowering both sides at same speed.

Putting Sections back Together

- Clean RTV and loose Kaowool from the steel sections and clean any bed material and rubbish.
- Make Kaowool blanket gasket to sit between refractory. Make this by cutting Kaowool blanket to the shape of the refractory so it sits over the width of the refractory and insulation but does not intrude into column. Tear the gasket so you end up with two gaskets the same shape but half the thickness.
- Cut a length of Kaowool rope long enough to fit round the inside edge of the steel flange.
- Fix the rope in place around the inside edge of the flange using RTV as glue. On the outside of the rope put a bead of RTV thick enough it will touch the flange above when tightened together.
- Attach D shackles to two opposing holes.
- Attach each of the block and tackle sets to the floor above and the D shackles.
- Lift the sections into place so they are in the correct orientation and the bolt holes line up.
- Once the section is close enough attach bolts and tighten them to take the weight. Remove the D shackles.
- Tighten bolts until the RTV touches both flange surfaces but do not tighten any further.
- Once the RTV has cured tighten the bolts fully and check for leaks by running air in the fluidised bed. Use soapy water to identify leaks.
- Refill beds with greywacke sand screened to 1 mm, bed material used in the gasifier is all less than 1 mm.

Gasifier Start-up Procedure

NOTE: In the start up procedure the gasification column is referred to as the BFB and the combustion column is referred to as the CFB.

Things to check before starting

- Pilot burners are operational. Check the ignition rod and flame detection rod are clear and not coated in char or tar. They should be sanded clean if they are not working.
- Remove the bung in the bottom of the siphon to check there is no wet sand blocking it. Use the piece of wire hanging beside the siphon to push up the pipe to check for wet sand. If it is blocked use the wire to work the wet sand free and clear the siphon. Once it is cleared dry sand will pore out of the hole. Remember to replace the bung when cleared.
- Remove the hopper on the bottom of the BFB cyclone to check it isn't full. Drain out any water and wash out any sand.
- Clean all the viewing ports at this stage so the gas flames can be seen clearly and monitored during the run. On the BFB and CFB make sure the valves are shut before removing the glass. Make sure the ports are put back in place tightly to ensure there are no leaks after removal.
- Check both O_2 sensors have the cable coming out the top and the bottom is open and clear. Remove BFB O_2 sensor line is clear, do this by removing the line at the union and using wire to clear away any residual char.
- All runs need to be scheduled on the white board in the control room. The gasifier can not be operated at the same time as the arc reactor or Fire Engineering hood so clashes need to be negotiated. So before running check:
    - Nano-tube arc reactor or fire engineering have no runs scheduled on white board. Running the arc simultaneously with the gasifier could cause an explosion risk if the gasifier leaks; the arc and power supplies will ignite any flammable gas.
    - Open duct valve to gasifier lab and lock in place with nut. The extractor fan can only extract from one lab at a time so check on white board in control room that fire engineering do not intend to use extractor that day.
- Turn on water supply to the pressure relief by opening valve on the wall in the back right hand corner of the lab. There is a ball cock in the reservoir at the top of the pressure relief which will regulate the height of the water.
- Turn on extractor fan labelled "Extract Fan E1" on control panel in the control room.
- Start extraction fans in room 175 with power interlock switch in main foyer.
- Once the louvers in the west wall have opened, start the "Fire Hood Fan" on the control panel and make sure it is at maximum speed using the "Fire Hood Fan Speed Control". The display should read approximately 67%.
- Check CO and O_2 sensors are operating by burning a match underneath then blowing it out. CO will rise when the match is blown out. O_2 will decrease when the match is bought under the sensor but increase again when it is blown out.

- Check H_2 sensor using a beaker with a small amount of acid and a piece of metal to generate gas. The reading will increase when the beaker is put under the sensor.
- Make sure the compressor is turned on in control room even if light is on so it won't turn off (the compressor can be started in other labs and could be switched off by someone else if not in control room). Check compressed air valve is open and water bulb is empty. To check air is in the line open the valve on the bottom of the bulb.
- After about two minutes the air line valve will open automatically.
- Check wood feed screw is tightly fastened and feeder valve is closed
- Switch on the power at the wall to gasifier control panel and also on the control panel, in the lab.
- Check clean up train lines are clear and take particular note of the valve as it blocks easily with high flow rates of gas and tar.
- Check that the controller is in the right mode and all faults are cleared; see controller operating instructions on how to do this.
- Start the temperature monitoring program run on the blower control computer. On the desk top when the computer is running appears an icon labelled "Gasifier Temperatures" Once open click on "Run" on the top tool bar and "Start".

Blower and Air Flow Start up

- Open the valve in the back left corner of the particle lab about 1/3 to allow pressure relief in the case the automatic valves shut off.
- Check the main red handled valve is fully open. Also ensure that two of the after-burner air supply valves and all other valves are open.
- Start the blower using the procedure hanging on the wall in the control room. Ensure the interface is switched on.
- Set the blower speed to approximately 15 Hz once it is running.
- The rotameters from left to right should be set as follows in Table 1.

Table 1 Rotameter air flow settings

Rotameters								
CFB Fluid	CFB Prim.	CFB Sec.	Siphon	Chute	BFB	A/B Flame	A/B Prim.	A/B Sec.
14	2	2	16	8	17	12	12	13

- Check the vortex flow meters and the orifice meter digital controllers are reading on the control panel. They should read similar values to those in Table 2 ±100.

Table 2 Digital Control Panel Flow Measurements

Digital		
BFB	CFB	A/B
300	300	15

- Check the pressure drop across both beds. The BFB should be ˜ 4 kPa and the CFB should be ˜ 7 kPa. If they are reading 2 kPa lower check the lines are not blocked by removing them and blowing on the end while checking for a change in the reading. If they are too high, there are several possibilities:

    - The air flow rates are too high, see Table 1 for details of what the rotameters should be reading.
    - The beds are too high, check by shutting down plant and removing the top port holes on both beds, then use the plum bob to measure how high up the columns the beds come. Mark on the outside of each column where the bed comes up to. The BFB bed should be about 500 mm from its bottom of the distributor and the CFB bed should be 200 mm from the bottom of its distributor. Sand can be removed by putting the bed into circulation mode and opening port in bottom of siphon.
    - The distributors have filled up with sand. In this case the plant needs to be shut down and the distributors removed following instructions in the Building guide.
- Take a set of all the readings and compare them to the standard set of start up data. Bed temperatures will read higher if the gasifier has been running recently.

LPG Gas Start up

- Red light on the switches on the control panel will come on when it is safe to start the LPG system. They do not mean the LPG is on or the switches.
- Check that all the LPG switches are turned off on the control panel. Up is off.
- Turn the gas supply to the room on using the key sitting on top of the electrical panel in the control room and make sure the top switch labelled Room 173 is turned on. The red light above the door into the lab will come on once the LPG is on.
- Check the gas pressure at the gauge sited on the inlet line on the wall above the rotameters, it should read ˜ 15 lb/in² (psi). If the pressure is low, check the valve is turned on, if it is then see the gas trouble shooting guide to solve the problem.
- Check the controller is set correctly and the main air flow to the afterburner is set between 12 and 14.
- Switch on the gas to the afterburner using the switch labelled A/B on the control panel.
- There should be a click after about ten seconds and a dull roar from the flame should be heard. View the A/B flame from below through the view port.
- If a flame does not start after ten seconds and the red lock out light is on wait 30 seconds and push the button to re start the gas and igniters. The lock out light is set back from the face of the controller.
- If the flame still doesn't ignite reduce the air flow a little. If it still doesn't ignite, see the gas system trouble shooting guide.
- Adjust the air supply to get a constant and stable blue coloured flame.
- Allow the A/B to burn for at least 3 minutes to make sure there are no flammables in the system before starting the BFB pilot burner. To get a stable flame with all burners operating the A/B needs to be hot. A good indication the A/B is hot enough to start the BFB pilot is the "Boiler Flue", temperature recorded by the blower controller computer, has reached a plateau on the plot.
- Check the switch on the BFB Gas control box is set to 0 and all the pilot burner gas line valves are open.
- Switch the BFB LPG supply on at the control panel.

- As long as the lock out button on BFB gas controller face is not lit, the pilot burner should ignite. If it is lit, push it in to start the burner. It should start after 10 seconds at which time the red light will come on.
- Check the pressure on the air regulator is set to about 2.5 bar.
- If the controller locks out adjust the air pressure down and retry, if problems persist, see the gas system trouble shooting guide.
- Using the view port on the top of the BFB check you can see the flame running. It can only just be seen but the glow is clear from above. Check the O_2 meter reading has dropped from above 20 % to about 15 % and the temperature above the BFB bed has risen from its initial temperature.
- Run the pilot burner for 3 minutes to make sure there are no flammables in the system before starting high pressure BFB gas supply.
- Check the flow adjusting valve on the high pressure BFB gas supply line is closed.
- Turn the BFB gas controller to 1 and an orange light will come on, on the controller. Check the O_2 meter is still reading 15 %.
- Slowly open the gas control valve and watch for a change in the O_2 levels, it can take up to 20 s for a change to occur from the time the valve is turned so adjust the gas flow slowly in small steps. Once the gas is igniting in the bed dull popping sounds will be heard but are of no concern, it is just LPG igniting above the bed.
- Adjust the gas rate until the O_2 is at 5 %. It will be hard to keep constant as the LPG burns as bubbles of air gas mixture break the bed surface, reducing the O_2 levels.
- Check the bed visually from above. There should be waves of orange and blue flames coming up through the bed and in the space above it.
- Allow the gas to the BFB bed to run on its own until the air supply to siphon is warm and the plugs in the bottom of the siphon are warm to touch. This will help stop any moisture clogging the bed again.
- Repeat the gas start up procedure for the CFB
    - Check the switch on the CFB gas control box is set to 0 and all the pilot burner gas line valves are open.
    - Switch the CFB LPG supply on at the control panel.
    - As long as the set back red button on the CFB gas controller face is not lit the pilot burner should ignite, if it is lit push it in to start the burner. It should start after 10 seconds at which time the red light will come on.
    - Check the pressure on the air regulator is set to about 2.0 bar.
    - If the controller locks out adjust the air pressure down and retry. If problems persist, see the gas system trouble shooting guide.
    - Once ignited increase the air regulator to just above 2 bar so there is sufficient air to cool the pilot burner
    - Using the view port on the side of the CFB check you can see the flame running. It can only just be seen but the glow is clear from above. Also check the O_2 meter reading has dropped from above 20 % to about 15 % and the temperature above the CFB bed has risen from its initial reading.

- Run the pilot burner for 3 minutes to make sure there are no flammables in the system before starting high pressure CFB gas supply.
- Check the gas flow adjusting valve on the high pressure CFB gas supply line is closed and air flows are still set to values in Table 1.
- Turn the CFB gas controller to 1 and an orange light will come on, on the controller. Check the O_2 meter is still reading about 15 %.
- Slowly open the gas control valve and watch for a change in the O_2 levels, it can take up to 20 s for a change to occur from the time the valve is turned so adjust the gas flow slowly in small steps. Once the gas is igniting in the bed dull popping sounds will be heard but are of no concern, it is just gas igniting above the bed.
- Adjust the gas rate until the O_2 is at 5 %. It will be hard to keep constant as the LPG burns as bubbles of air gas mixture break the bed surface, reducing the O_2 levels.
- Check the bed visually from the side. There should be waves of orange and blue flames burning at the end of the port and bed material should be seen being blown about, it will appear dark at this stage.

General Heat up

- All readings and measurements need to be constantly monitored through-out the heat up and running period. The record sheets need to be filled out every 30 mins while the gasifier is in operation and any valve adjustments or changes made must be noted on these sheets along with the time of the change.
- Beds and flames must also be examined visually every 30 min through the viewing ports on the top of the BFB, through the side of the CFB, through the bottom of the A/B and through the side of the wood feed hopper. Record notes of the state of each so change is noticed.
- Examine joins in pipes and columns both visually and by touch to check for leaks. If you are not sure about an area, check it out by brushing it with soapy water and look for growing bubbles. <u>If any leaks are detected, shut down the plant immediately.</u>
- The auto-ignition temperature of the LPG is about 640 °C. At this temperature the beds become much quieter and popping stops. The temperature above the BFB will begin to drop but this is no cause for alarm as the gas is igniting in the bed, it will end up below the temperature of the bed and will rise as the bed temperature rises from here.
- The bed material will appear orange and the waves of flame above the BFB will no longer be seen above auto-ignition temperature.
- At this temperature air rates should be increased to start bed circulation and heat syphon etc, the rotameter settings should be as follows in Table 3.

Table 3 Rotameter Settings for Bed Circulation

Rotameters								
CFB Fluid	CFB Prim.	CFB Sec.	Siphon	Chute	BFB	A/B Flame	A/B Prim.	A/B Sec.
16	7	6	18	10	17	12	12	12

- Check bed material is circulating by visual inspection through the BFB viewing port. When bed material is circulating waves of dark red material will be seen falling from the siphon entry into the hotter orange BFB bed.
- Once bed material is circulating the gas rates can be increased to reduce the oxygen content of the combustion gases. Slowly increase the high pressure gas rate to each bed until the O_2 reading drops to 2 %.
- Take particular care at this point to note get the circulation rate too high for the syphon to keep up and end up having bed material come out the heat exchangers into the sand catcher. Check it by opening the door on the bottom, use a metal bucket to catch the bed material. Be careful as the sand catcher and any bed material in it will be very hot.

Wood Feeder Start up

- Turn on nitrogen cylinder and tighten regulator down to 200 kPa. Adjust flowrate to 12 L/min.
- Turn off main BFB LPG supply by turning the controller switch to 0 and closing ball valve but keep the pilot running until steam is started to ensure no combustibles. Keep an eye on the O_2 levels at this time to make sure they do not reach 0.
- Make note of the pressure drop across the BFB and take a full set of readings.
- Increase after burner air supplies to the following settings.

Table 4 shows the A/B air flow rates required once the steam started

A/B Flame	A/B Main	A/B Dilution
12	18	18

- Switch on the steam on the front of the control panel.
- Turn on all electrical switches but ensure VSD is set to 0. Make sure switch on auger is set to FWD.
- Slowly increase auger speed to 2. Monitor the BFB O_2 and temperature as you do this. The temperature should drop as should the O_2.
- Once wood feeder is set change the O_2 measurement from the BFB to the A/B. Adjust the air flow rates to the A/B until the O_2 is about 17 % and the maintain the temperature below 1000 °C. The temperature is more important so do not worry if the O_2 does not 17 %.
- The after burner may begin to roar at this stage but this is nothing to be concerned about, it is due to a change in the fluid dynamics at high temperatures and velocities.

Steam Start up

- Once the “Boiler Flue” temperature plot has reached a plateau steam can be introduce to the BFB.
- Check air feed temperatures in the siphon, chute and BFB are all above 120°C.
- Open the BFB water control valve until rotameter begins to rise above 0. Watch the pressure drop across the BFB bed as it does and reduce the air flow to maintain it at its initial value.

- Increase the water flow rate in increments turning down the air flow as you do so to maintain the pressure drop across the bed until the air flow is completely shut off. At this stage the rotameter will be reading about 80 L/h. Watch the temperature of the distributor does not drop below 100 °C as the steam will condense if it does.
- Increase the siphon and chute water supplies incrementally, reducing the air supplies as you go. There is no reading to check the flow through each is the same as with the air but the chute rotameter should read 5 mL/min and the siphon should read 5 mL/min. Watch the temperatures do not drop below 100 °C during this, as the steam will condense and they will fill up with water.
- Check the BFB viewing port to ensure bed material is still circulating.
- The temperature of the BFB will have dropped during this time, to bring it back to operating temperature of 800 °C increase the air rates and gas rate to the CFB.
- Turn off BFB pilot burner and at control panel.

Cleaning Train Preparation and Operation

Product Gas Conditioning Train Description

Dirty product gas direct from the gasification column contains tars and particulate matter which need removed before the composition can be can be measured using gas chromatography (GC). The clean train removes the tars and particulates in a series of particulate filters and wash bottles based on the European Union Guideline for Sampling and Analysis of Tar and Particles in Biomass Producer Gases (Simell, Stahlberg et al. 2000; Neeft, Knoef et al. 2002). Tars are sampled from the train using the solid phase extraction method (Brage, Yu et al. 1997) using amino phase columns and analysed using high pressure liquid chromatography (HPLC) with a method developed especially. A schematic of the clean up train is shown in Figure 1.

The in take to the product gas clean up and analysis train is situated at the exit of the product gas cyclone. The product gases then pass through a particulate filter to remove entrained bed material and char from gasification column. The filter temperature is maintained above 350 °C to prevent tars condensing on the surfaces, provision was made to trace heat the filter but the hot producer gas insulate in the tube was sufficient to maintain the temperature. Quartz thimble filters (26 mm ID x 60 mm long) were purchased from Scientific Supplies (product number QF-20).

Directly after the particulate filter is, a T where amino solid phase absorption columns are connected into the line to take gas samples to analyse tars.. The gas then passes through six wash bottles connected in series to remove any remaining particulates, the tars and water vapour. For obvious reasons the Micro GC can not take particulate matter or tars. The water vapour must also be removed as it gets held up in the Micro GC column pores. The wash bottles are connected in series with the temperature in the first four being maintained at about 15 °C by a water bath with continuous flow through it. The wash bottles in the water bath are half filled with clean industrial grade iso-propanol in which the tar compounds are almost totally soluble (Neeft,

Knoef et al. 2002). The last two was wash bottles temperature is kept very low by keeping them in a mixture of acetone and dry ice. These wash bottles are filled with glass wool to increase residence time and condense out any remaining water, tar and solvent.

The gas samples are sucked through the system with a diaphragm pump which is situated after the wash bottles as it can only operate with clean dry gas. The pump is a Capex V2 12 v DC diaphragm pump connected with power supply and 12 v variable voltage output for varying the pump speed. The pump can pump up to 3 l/min at 2 bar_g but in this application is restricted to about 1.5 l/min or 90 l/h. The gas it then goes through a rotameter measuring the flow rate and there is provision for a gas volume totaliser but it is not installed at this stage. The gas totaliser would allow the wash bottles to be used for quantitative tar analysis if installed. The Micro GC has a entrained liquid and particulate genie on the front end for protection so the system down stream of the pump is kept at a positive pressure of 10 psi to maintain keep this operational. The pressure is monitored manually and adjusted using a needle valve down stream of the Micro GC supply point.

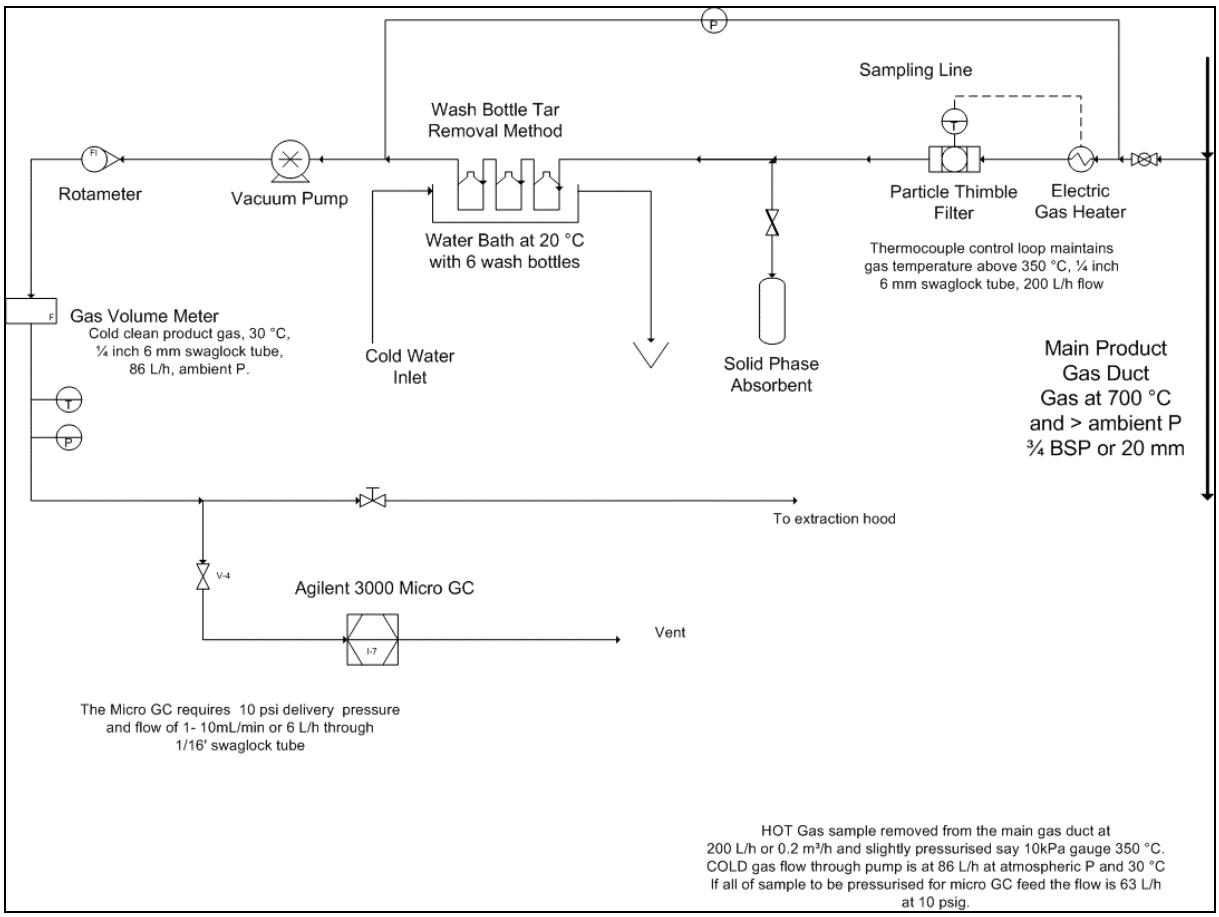


Figure 1 Product gas cleaning and gas sampling train.

Set-up Before Operation

- Clear all lines of built up tars and particulates with iso-propanol or dichloromethane solvent. Compressed air can be used to clear heavily fouled components. The valve may need to be disassembled if it is blocked.
- The particulate filter does not have to included in the train but in the case that it is follow these steps:
    - Weigh the filter and record the weight
    - Insert the filter cartridge over the sleeve on the inside of the main holder. Hold in place with lacing wire.
    - Screw the assembly back together with adequate thread tape on the thread.

- Fill four of the six wash bottle half full of iso-propanol (IPA) solvent. This is stored in the flammable goods safe in the particle analysis lab next door.
- Fill the remaining two wash bottles with glass wool
- Mix dry ice (available from Trevor Berry or Glenn Wilson) in a one to one by weight with acetone (also kept in the flammable goods safe in the particle analysis lab next door) so there is enough to half fill the two vacuum cylinders used to hold two of the wash bottles.
- Attach the wash bottles in order so there are four wash bottles in the water bath. The first three should contain IPA and the last one should contain glass wool. Make sure the wash bottle heads are attached around the right way so the gas comes in via the glass tube through the solvent and exits out the top.
- Put the remaining two wash bottles into the dry ice acetone mixture so the gas flows through the IPA containing bottle first. Again check the gas flows into the wash bottle down through the tube and out via the top.
- The Swaglok tube fittings should be attached together so they are in the order shown in Figure 1 above.
- Attach PVC tube from final wash bottle head to the pump inlet
- Make sure all the ground glass fittings are well attached and check for leaks by running the pump for a short burst. If there are no bubbles in one of the solvent containing wash bottles then there must be a leak down stream between it and the next bottle which is bubbling. The connection between the wash bottles and heads is most likely to leak so push the head and bottle together while slightly twisting to check otherwise check the connections between the heads.
- Make sure all lines between pressure side of pump and flue are connected in correct order and swaglok fittings are tightened.

Operation

- Open the valve down stream of the micro GC completely so gas can be pumped through unimpeded.
- Open the main shut off valve at sample point.
- Switch pump on at wall
- Switch on pump at the variable voltage speed controller and increase current to about 1 A.
- Close valve in the control room until the pressure of the gas fed to the micro GC is between 10 and 15 psi. Because a diaphragm pump is used to circulate the gas the reading jumps around a lot so a visual estimate is sufficient for the pressure.
- Now gas is being continuously pumped start the micro GC to periodically taek samples.
- The pressure of the gas being fed to the micro GC should be checked periodically along with the level in the wash bottles.
- The high moisture content of the product gas means they fill up quickly so this needs to be checked regularly. When the wash bottles are full the pump should be stopped and the main valve closed and the wash bottles should be emptied into a labelled container for appropriate disposal. The wash bottles are then refilled as at start up.

Micro GC Preparation and Operation

Background

In April 2005 an Agilent 3000A Micro Gas Chromatograph was purchased from the New Zealand agent for Agilent, Biolab, to analyse samples of product gas from the biomass gasifier. Colin Welkin is the representative of Biolab who recommended the system that was purchased and provided much needed support and training after installation. The chromatograph is set up with two channels fitted with a Molecular sieve and Plot Q columns for separating the different components that make up the product gas. An on board gas cylinder allows carrier gas for one channel to be stored in the machine. The channel configurations are summarised in Table 5. The Plot Q column is fitted with a back flush valve which opens after the analytes of interest have passed through the pre-column to flush unwanted analytes back through the injector. This keeps the analytical column clean and allows it to hold a calibration longer. Both columns are fitted with single filament thermal conductivity detectors which measure the thermal conductivity of the analytes relative to the carrier gas. Both columns also have heated inlets.

Table 5. Micro GC configuration.

Channel	**Channel A**	**Channel B**
Column	Molecular Sieve	Plot Q
Carrier Gas	Argon	Helium
Injector	Backflush	Fixed volume
Detector	Single Filament Thermal Conductivity	Single Filament Thermal Conductivity

Instrument grade helium and argon carrier gases were purchased from BOC gases. The compositions of these are shown below in Table 6.

Table 6. Composition of carrier gases as supplied by the supplier, BOC, on delivery.

Helium		**Argon**	
Component	**Volume %**	**Component**	**Volume %**
Helium	>99.99	Argon	<99.99
Argon	<5 ppm	Carbon Dioxide	<1 ppm
Carbon Dioxide	<5 ppm	Carbon Monoxide	< 0.5 ppm
Carbon Monoxide	<5 ppm	Hydrocarbons (as Ethane)	< 0.5 ppm
Hydrocarbons (as Ethane)	<1 ppm	Hydrocarbons (as Methane)	< 0.5 ppm
Hydrocarbons (as Methane)	<1 ppm	Moisture	< 5 ppm
Moisture	<10 ppm	Nitrogen	< 5 ppm
Nitrogen	<50 ppm	Oxygen	< 5 ppm
Oxygen	<10 ppm		

The operating conditions are shown in Table 7; initial values for these were taken from Agilent test set up paper on GC. The operating conditions were changed by trial and error to get the best separation for the gas composition measured from the FICFB gasifier system. The chromatograph was set by first thermally cleaning the columns at

290 °C for at least one hour until a regular flat base line was obtained, baking out unwanted analytes captured in previous operation. The chromatograph is calibrated before each use with 3 separate samples of beta standard calibration gas. The GC was calibrated using a beta standard specification gas mixture made up by BOC gases. The beta standards are calibrated to ±0.5 vol.% uncertainty for the most inaccurate component. The composition and uncertainty of the calibration gas is shown in Table 8.

Table 7. Micro GC operating conditions

Condition	Channel A	Channel B
Sample Inlet Temperature	95 °C	95 °C
Injector Temperature	95 °C	55 °C
Column Temperature	110 °C	60 °C
Sampling Time	15 s	15 s
Injector Time	10 ms	15 ms
Run Time	180 s	240 s
Column Pressure	207 kPa	138 kPa
Backflush Time	15 s	0

Table 8. Composition of beta standard used to calibrate chromatograph.

Component	Volume %	Uncertainty (± Vol. %)
Hydrogen	44	2
Carbon Dioxide	27	1
Carbon Monoxide	11.0	0.4
Methane	10.5	0.4
Nitrogen	3.1	0.1
Ethane	0.55	0.02
Ethene	3.7	0.1

Operation of Micro GC

The micro GC is a very complex piece of equipment which requires much more tuition than just reading this guide. There is a book on chromatography that explains the basics for beginners called “The essence of chromatography” by Colin F. Poole. It is available from the Science library. This will give an overview of the theory behind chromatography and how the gases are separated and what determines the elution rates and sample times and how parameters affect each other. There is also a guide written especially for the Agilent 3000 micro GC. It is stored in PDF form on the Compaq lap top used to operate the Micro GC. The program which runs the Micro GC called Agilent Cerity QA-QC contains its own self help program and also tutorials on how to use the program.

- Things to remember before turning GC on:
    - On board cylinder filled with helium has a pressure greater than 900 psi. Check the swaglok link between the on board cylinder and channel B are tightened and linked to the correct ports.
    - Argon cylinder has pressure in it and lines between cylinder and channel 2 connection on the back of the GC are connected through the filter and swagloks are all tightened.
    - Network cable linking the Micro GC to the laptop is plugged in.

- Turn on power to the GC and Laptop. On the laptop open “explorer” and connect to the Micro GC. Check the channels are configured to the correct carrier gas and set up.
- Open the program Agilent Cerity QA QC. Load the desired method and allow the Micro GC to warm up. If there is no suitable method use the tutor installed in the program to help set up a method and use the values in Table 7. Micro GC operating conditions as a starting point to develop a suitable method.
- Run the Micro GC until a consistently flat base line is formed on both channels.
- Make sure sample pump is off before disconnecting the sampling line. Connect the calibration gas to the genie on the front of the Micro GC and set the pressure on the regulator to 15 psi.
- Take several samples until the area under each curve is consistent to about 1%. If the separation is not acceptable use the tutor to alter the method to get the desired separation.
- Use the area of each of the gases to calibrate the method for the percentage the gas is calibrated for.
- Disconnect the calibration gas and reconnect the sampling train system.
- Using the newly calibrated method sample the product gas from the gasifier clean up train.

Moisture Content Measurements

Product gas moisture content was assessed by taking samples of the gas stream at a port above the cyclone on the exit of the gasification column. The moisture is condensed out of the gas stream in a cold trap. A short section of stainless steel tube insulated with ceramic fibre leads the gas into a 100 mL cold trap filled with glass wool (available from dry chemicals store).

- The cold trap sits inside a thermos filled with a mixture of acetone and solid carbon dioxide (ratio of 1:1 by weight) which has a temperature of -77 °C (Merck 1986). See Trevor Berry for dry ice and acetone.
- The sample is pulled through the cold trap by emptying five litres of water from a glass vessel.
- Heat the cold trap and upstream pipe work are weighed at 105 °C in oven for 20 min to drive off the moisture.
- Weighed again to find the mass of moisture. The condensed tars are then washed out of the cold trap with DCM solvent, use pipe cleaners to clean tubes and dispose of the waste appropriately with chlorine waste.

Tar Measurements

The solid phase extraction (SPE) columns contain 100 mg of amino-propyl silane bonded silica gel, which tars are absorbed on to, are commercially available from JT Baker (product number 7088-03). The solvents used are dichloro-methane (DCM) and isopropanol (IPA), both instrument grade with a purity of 98 % or greater. 1 mL syringes for preparing SPE columns, 50 mL syringe for pulling the samples through

the columns and luer lock fittings to connect syringe into column and column to duct work. Needles are required to connect the columns into main sample line.

Solid Phase Extraction Capture Procedure and Sample Preparation

- Prior to sampling the columns are pre-treated by flushing them with 0.5 mL of DCM
- Heat columns to 100 °C for 5 minutes in an oven.
- Once the columns are pre-treated the covered needle is attached to one end and a rubber stopper inserted in the other to prevent contamination in the atmosphere until the column is used.
- Attache luer lock needle to the end of the SPE column so it can be inserted into the $1/16^{th}$ swaglok tube fitting, in the sample line.
- Connect 50 mL syringe connected to a luer lock 3-way valve is inserted in the other end of the SPE column to pull a 100 mL sample through the column.
- Samples are taken by inserting the needle into the sample line and pulling a sample through the column with the attached 50 mL syringe. To get a 100 mL sample the syringe is emptied using the 3-way valve and another sample is pulled through the column with the syringe.
- Immediately seal the needle and the end sealed with a rubber stopper. The needle is left attached so that any tars that condense inside it during sampling can be analysed giving a more accurate sample. The samples are then analysed as soon as possible using the HPLC method developed by (Devenish, 2006).

Gasifier Shut Down

Stopping Gasification

- Stop wood flow by turning off VSD at wall
- Close wood feeder valve.
- With the O_2 sensor monitoring the A/B O_2 reduce the main air rate as gasification stops to stop the pilot going out. Try and keep the O_2 at about 16 %. If it does go out you will not be able to restart the after burner for at least 2 minutes as it locks out. If the flame goes out this is a sign gasification has stopped and only small amounts of combustable gas are being produced so continue with the procedure by increasing the BFB air rate slowly as this will help burn out any remaining char.
- Incrementally reduce the steam flow rate while increasing the air rate through the BFB to maintain a constant pressure drop across the bed until water fed to the boiler is totally stopped.
- Repeat above process with water feed to chute and syphon.
- Leave air running through bubbling bed until O_2 reaches at least 16 %. This is a sign that the remaining char is all burnt out.
- Follow shut down procedures for other parts of the plant as follows.

LPG Gas Shut Down

- This is under taken with the plant in its normal running state with all air flows and extraction systems still running as per normal.
- Shut down procedure is the same for both the CFB and BFB columns.The order in which the main burners of the CFB and BFB columns are shut down does not matter, but the after burner should be left running while these are both shut down.
- Shut off the high pressure gas supply at the flow control valve.
- Check the O_2 reading rises to about 15 % then turn gas controller switch to position 0. The orange light on the controller should go out but the red light should stay on.
- Leave the pilot burners running for at least 5 min with air flow through the bed to make sure all flammables are burnt out.
- Turn off both switches at the main control panel to cut the gas flow.
- Allow the A/B to run for a further 5 min for safety reasons. NB. Air supplies to the A/B will need to be reduced once gasification has stopped to stop the A/B pilot burner going out.
- Switch A/B off at the control panel but leave air running through both beds and A/B.
- Switch off gas in control room and remove key and put back on top of the control box.

Plant Shut Down

- Check primary and secondary air supplies on the CFB are pointed down so no sand flows back into them.
- Reduce air flows to those in Table 1. Increase air flow through the A/B to allow for pressure increase in blower
- Turn off Nitrogen Purge
- Stop blower
- Turn off extraction hood, this will cause the controller to close all the air supply valves.
- Once extraction system speed reaches 0, the extraction duct labelled “Extract Fan E1” can be turned off.
- The blower computer should be shut down in accordance with the procedure outlined on the poster in the control room.
- Close extraction hood valve so air flow open to the fire engineering position and fix in place with nut.

Operations Check Lists

The following table can be use to check off tasks required during gasifier operation. The check lists are saved as an excel spreadsheet in Jocks work in:
Post Graduate\Gasifier; Thesis\Results\Gasifier Measurements.xls

Gasifier Check List

	Initial								
	Date								

Pre-run checks

Pilot burner ignitor and flame rods checked									
Controller Power on									
Controller in normal mode and faults cleared									
Syphon clear									
BFB cyclone hopper empty									
Viewing ports all cleared									
O2 sensors right side up									
BFB O2 Sensor clear, no char in line									
Schedule clear, clashes sorted									
Extractor duct open to gasifier lab									
Arc reactor off									
"Extractor Fan E1" on									
"Fire Hood Fan" on									
Particle Lab Fans on									
CO and O2 sensors have current calibration									
H2 sensor has current calibration									
Compressor switched on, valve open, water empty									
Pressure relief water valve on									
Wood feeder full									
Screw feeder connected, valve closed									
Check Clean up train valve will open and is not blocked									
Start temperature recording program on blower computer									

Air start-up

Valve in particle lab 1/3 open									
Main valve in lab fully open									
Blower speed ~15 Hz									
Correct rotameter settings									
Bed pressure drops correct									
Bed heights reasonable									

LPG Start-up

Red lights on control panel on, switches off
LPG on in control room to 173
LPG gauge reads 15 psig
A/B pilot air 12 to 14
A/B controller set to 1
A/B LPG on (red light)
Check flame visually, temp.
A/B "Boiler Flue" temperature plateau
BFB controller set to 0
Trun on at main controller
Check O2, temp, visually
Main LPG Ball & needle valves closed
Controller to 1
open ball then needle slowly
adjust needle until O2 ~5 %
Repeat with CFB

General Heat up

All readings monitored constaly and recorded every 30 min
Visual inspections every 30 min
Check for leaks
Increase air rates at 680 °C
Increase LPG and check O2

Wood Feeder Start up

Nitrogen Cylinder on

Switches on in Fwd mode
Slowly increase screw speed
Check O2

Steam Start up

Temperature ~850 °C
Increase A/B air rates
Main BFB LPG off
Check BFB distributor, syphon and chute temps > 120
Steam on at control panel
Increase water decrease air
Keep Bed ?P constant
View BFB, Bed material circulating
BFB pilot off at main control panel

Stopping Gasification

VSD off at wall
close wood valve
Monitor A/B O2 and adjust air accordingly
Reduce steam flow, increase air to BFB, chute, syphon

LPG shut down

Shut off main LPG supplies at valve and control panel
Monitor O2 then shut off pilot burners at main control panel
Leave A/B on for at least 5 min
A/B off at control Panel
LPG off in control room

Plant Shut Down

Air supplies pointed down
Reduce air flows, increase in A/B
Nitrogen Purge off
Stop Blower
Extraction hood off
Extract fans off
Compressor off
Blower and Micro GC computer off
Turn off pressure relief water supply

Measurement Record Sheets

Every 30 minutes during operation of the gasifier the temperatures, flowrates, oxygen concentrations, need to be recorded for safety so changes from normal operation are note. These measurements can also be used during calculations. The record sheets are saved as an excel spreadsheet in Jocks work in Post Graduate\Gasifier; Thesis\Results\Gasifier Measurements.xls

Table 9. Flow measurements for air supplies, extraction system, water supplies to boilers, nitrogen purge and LPG to CFB.

Flow Readings

Date

Time	Rotameters (cm)									Digital					Water Rotameters			Nitrogen	LPG
	CFB Fluid	CFB Prim.	CFB Sec.	Siphon	Chute	BFB	A/B Flame	A/B Main	A/B Dilutio	BFB	CFB	A/B	Fire Hood	Blower	BFB	Chute	Siphon	Purge	Flowrate

Table 10. Temperature measurements for the two columns and after burner.

Temperature Readings														
Date														
Time	CFB Temperatures [°C]								BFB Temperatures [°C]				A/B	
	Distributor (1)	Bed	Prim/Sec inlet (2)	Op. pilot (3)	Top (4)	Prim Air	Sec Air	Cyclone	Distributor	Bed	Above bed	Cyclone	A/B	A/B Outlet

Table 11. Temperature measurments for miscellaneous areas such a siphon, clean up train, extraction duct and steam supplies.

Temperatures 2								
Date								
Time	Miscellaneous		Tar Train		Wood Feeder		Air/Steam supplies	
	Duct	Siphon	Before	After	Top	Hopper	Syphon	Chute

Table 12. Measurements oxygen, pressure drops, gases levels and wood feed.

Other Readings														
Date														
Time	O_2 [vol. %]		Bed ?P [kPa]				Gas Pilot Air Flows		_PG P [psi	Gas Levels		Wood Feed		Pressure Relief
	CFB	BFB	CFB	Stability	BFB	Stability	CFB (14)	BFB (10)		CO [ppm]	H_2 [%LEL]	VSD settin	opper Lev	

Material and Energy Balance Use

The heat balance was primarily set up to calculate the bed material circulation rate under gasification conditions; however, it is useful for calculating the char circulation rate and getting the data into a useful format. The balances uses the product gas composition (from the micro GC) and the moisture content of the product gases to calculate a heat of formation and sensible heat required to form the gas. The heat required for the reactions is assumed to be all supplied to the gasification column in hot bed material circulating from the combustion column. The energy balance uses solver to find the bed material circulation rate supply the required heat to generate the product gas. The char circulation rate is calculated uses a mass balance on the combustion column. The oxygen concentration and LPG feed rate are recorded and oxygen not combusted by the LPG is assumed to be combusted by char circulated from the gasification column. The balance works by using solver to find the char circulated to balance the oxygen.

The excel file for the heat balance can be found in Jocks work under Post Graduate\Gasifier; Thesis\GC and Gas Analysis\Heat Balance.xls. Make sure when opening the file that the you click to enable macros, because the balance is run by a VBA program. To run the balance:

- Open the work sheet labelled “Measurements”.
- Enter the date and time in the top corner of the worksheet as this is the only way to determine between runs.
- Enter data recorded in the measurement record sheets in the rows labelled “Gasification” under the relevant heading. The headings are in the same order as the measurement record sheets.
- Enter data from the micro GC analysis and moisture measurements in the relevant cells.
- Click the big orange button labelled gasification.
- Solver will take over and ask you if you wish to accept the values found. Click yes unless it has an error, then there will be some debugging required. It is most likely in the case of an error that one of the temperatures or flowrates are inserted incorrectly or there was an error in the measurement so check all measurements make sense before changing anything in the spreadsheet.
- You will then be looking at the results page with all the data neatly recorded in the relevant columns.
- If there is a mistake in one of the values entered the best way is to delete the relevant row in the results sheet (make sure all data in the row is deleted) and re-enter the values and run the program again.

Gas System Trouble Shooting Guide

Gas Burner Set up Checks

- Outer tube should be square to centre unit and both are sealed with Kaowool paper between them.
- LPG tube is central in assembly
- Flame rod is not touching anything in areas where it is not insulated
- Spark electrode is not too close to wall or other tubes and spark gap to LPG tube is approximately 1mm

Gas Pressure Low

If the gas pressure on the gauge above the rotameters drops below 150 kPa, check the following things in order:

- Gas is turned in the control room.
- Main gas supply line valve on back wall is turned on.
- Indicator on main regulator for low pressure gas to pilot burners. If indicator is green there is no supply to the pilot burners. Back pressure in low pressure line has got too high and caused it to vent. Regulator needs to be reset.
- Solenoid valve in the Semi-Scale lab, this cuts out when ever the power goes out. It is situated at the far end beside Dave Browns office. If unsure get help from one of the technical staff.
- The pressure of the gas cylinders beside the roller door into the semi-scale lab. The key to get into the cage to do this is available from in the technicians' workshop.
- If the gas pressure here is low the gas bottles need to be changed over and new ones ordered. See one of the technicians about this.

Pilot Burners Fails to Ignite

If when starting the pilot burners and they do not ignite and lock themselves out there are several simple problems that may have occurred. So before doing anything else check the following:

- Gas supply pressure is 150 kPa, if not follow the procedure above to find the problem.
- Air pressure of compressed air lines.
- All gas and air line valves are open.
- Check there is a good connection with both igniter and ionisation sensor leads and their probes.
- Check solenoid valves have opened by checking pressures of both. If they have not opened:
    - Check there is no back pressure on the valve at pressure measurement points.
    - Check solenoid is sitting correctly on the valve and the connections are good.

Pilot Burner Ignites then Goes Out

In the case that the flame ignites then goes out the micro ammeter can tell us if we have enough current flowing for the controller to know if there is a flame or not. Sand in the bed can put the flame out as can a change in the flame pattern at a certain temperature; this can be seen in the reading on the ammeter.

To connect the ammeter (Figure 2) into the circuit, the lead wire to the ionisation sensor should be removed from probe at the connector strip (see Figure 1). **THE AMMETER IS TO BE USED ONLY IN CASES FOR TROUBLE SHOOTING THE PILOT FLAME NOT IGNITING AND SHOULD NEVER BE USED DURING GASIFIER OPERATION.** The ammeter can then be connected into the circuit with the red wire connecting to the lead of the probe, the black connecting to the probe itself and the green wire connected to earth (all of the gasifier structure is earthed). Make sure they are all metal to metal connections as the currents are very small and can easily be interrupted with a bad connection.

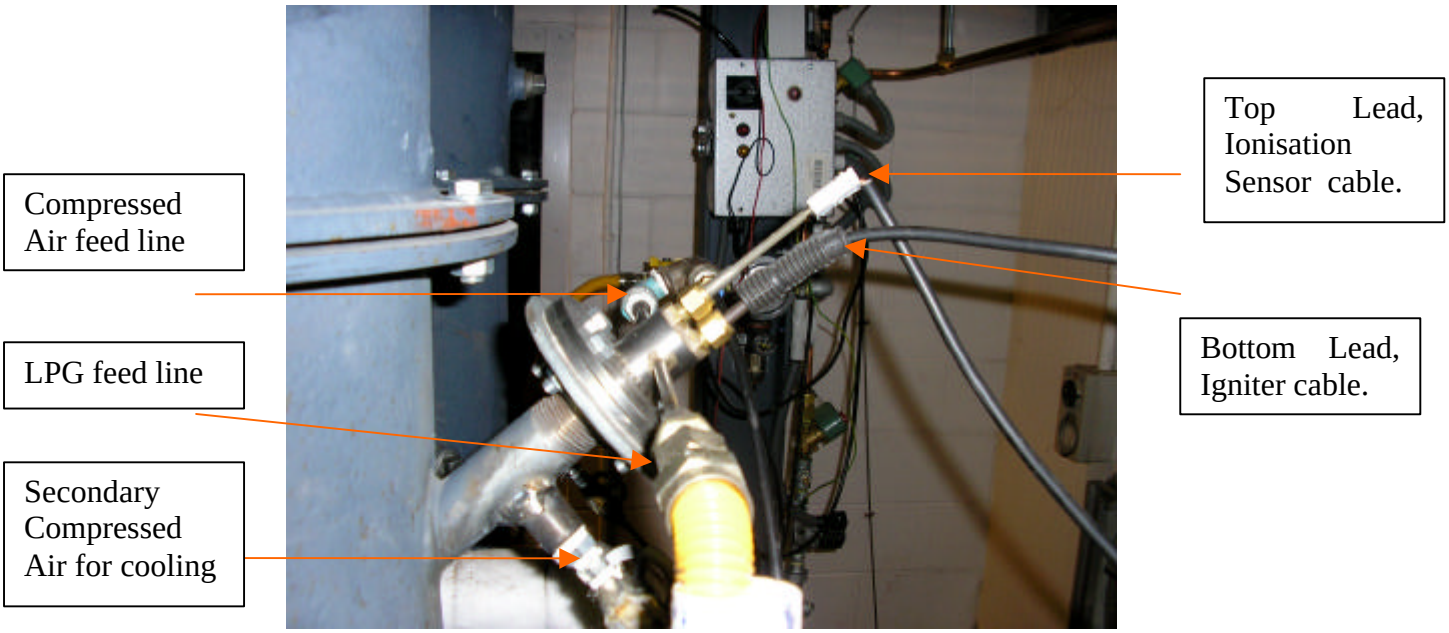


Figure 2 Pilot Burner External View

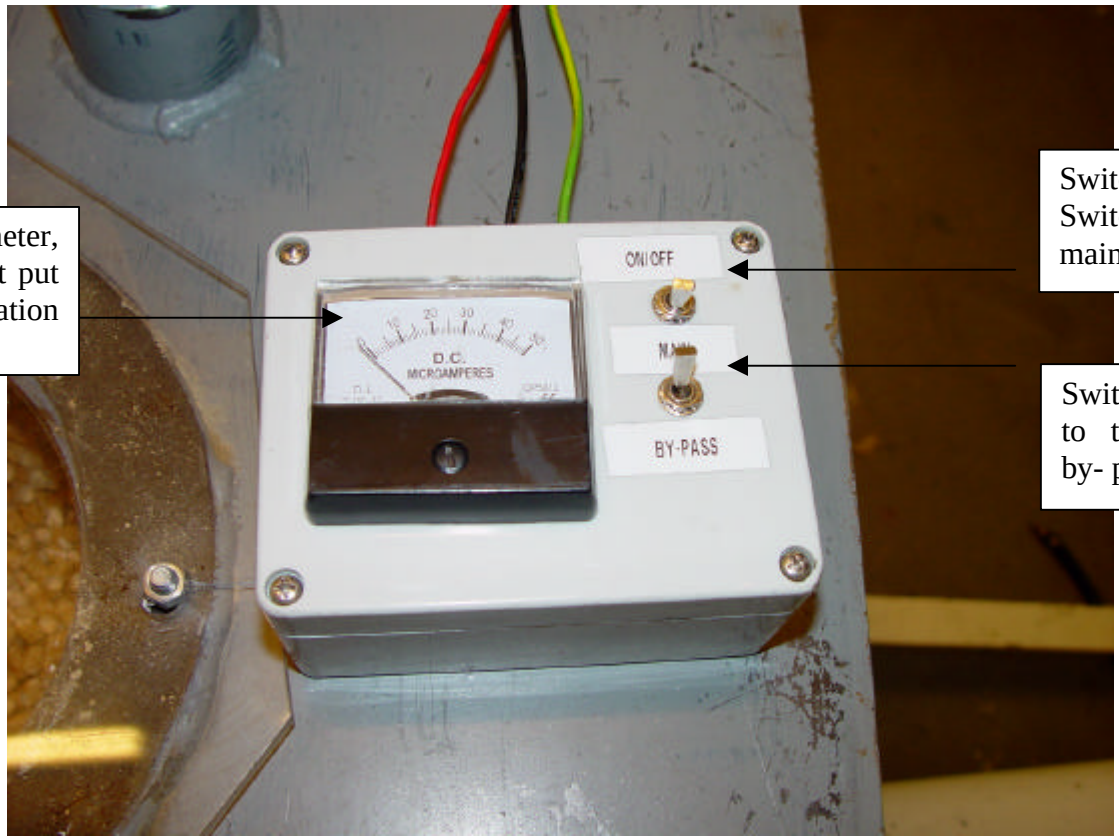


Figure 3 Micro Ammeter and Ionisation Senor by-pass

- To get a reading on the ammeter to check the current flow Switch 1 should be down and Switch 2 should be up.
- Once the micro ammeter is connected in the circuit, restart the burner and watch the reading, it should be above 20 µA to stop it going out under violent conditions.
- If the reading is less than 20 µA and is not stable it is likely that the combustion is not good enough, so one of the following can be tried:
    - Compressed air pressure is too low or too high. In this case adjust to the compressed air regulator to get a better reading, but be aware that this may increase the flame temperature and so the secondary may also need adjustment.
    - Secondary air flow rate is too high. Reduce the secondary air supply to see if it makes the reading more stable.
    - LPG rate is too high/low. This can be adjusted at the regulator, although this will mean the air and secondary air rates will also need adjustment. It is suggested this is not tried unless nothing else has worked and you have to remove the whole burner, because the flame pattern may change dramatically and should be observed.

If stable flame with a satisfactory sensor reading is not achieved then the blower air supply should be shut down following the procedure above and LPG should be turned off at the control panel and the whole pilot burner should be removed from its mounting. Once out of its hole the outer tube should be removed and the sensor and inside of the casing tube need to be inspected for fouling that may stop current flowing. Any major fouling needs to be sanded off so the metal surface can be seen. The following can then be tried to fix the problem:

- Reassemble the burner outside its mounting using the nuts, try to start it and observe the flame produced. The two traits we are looking for in our flame are:
    - High gas velocity so bed material can not get up the tube to put out the flame.
    - Flame burning back in tube so we get a good sensor reading.
    - However we don't want the flame getting too hot or burning too far back up the tube as it will heat up the tube and it will corrode. In extreme cases if the flame is too hot it can cause the sand to fuse.
- If flame burning too far back up tube adjust the gas and air rates until a better flame pattern is found.
- Once a stable flame pattern is found note the setting and check the flame temperature. If it is too high the tube will glow red hot, in this case either the primary air rate needs to be increased or the secondary air needs to be increased.
- Check the burner can start under the new conditions and reassemble it.
- The ionisation sensor can also be moved, bent and repositioned to reach across the flame more if the reading is still unsatisfactory.
- If none of these work the length of the air and gas lines may need to be adjusted but expert help is required to do this.

Construction

Cast Refractory Sections

The gasifier was originally design to be constructed entirely from cast sections of insulated refractory housed in steel cylinders bolted together. It soon became apparent that it was more favourable to have the more complex sections constructed from stainless steel to allow much easier construction and reduced heat up time without compromising heat losses and other differences from a real system. The cast sections consist of four separate layers, the rolled steel outer, calcium silicate board backup insulation, vermiculite insulation and the refractory hot face.

The steel sections are cylindrical in shape with an inside diameter of 390 mm and stand 495 mm tall and are constructed from 4 mm mild steel. Each end of the sections has an 8 mm thick flange which over hang the inside of the sections 20 mm with eight evenly spaced 12 mm holes for M12 bolts. The inside of the sections are lined with Calcium Silicate board, on the gasification column the board is 20 mm thick and 50 mm thick in the combustion column which has a smaller internal diameter. The boards are cut to shape and glued to the inner surface of the steel sections with builders glue (Selley's no more nails).

The inside of the calcium silicate board is lined with vermiculite insulation cast on the inside of the calcium silicate board. The vermiculite is mixed with calcium aluminate cement (Electroland brand) in a ratio of seven parts vermiculite to one part cement, which is about one to one on a weight basis. Water is added at a rate of 2 – 2.5 times the weight of the vermiculite and cement mixture to get a consistent mixture which forms balls that hold their shape when compressed. The steel sections have steel rods welded to the inside to hold the vermiculite in place once dried. The inner surface is cast with cardboard tubes. The insulation is then left to dry for 48 hours in a warm dry room. In the both the gasification and combustion columns the vermiculite insulation layer is about 40 mm thick.

The hot face refractory lines inner most surface of the columns where it with stands the abrasion and temperature of the fluid beds. The refractory is also cast in place with cardboard tubes. The refractory cement is a Sila product, silicast LC 180 which is a 93 % alumina low cement castable product suitable for temperatures up to 1800 °C. This product is over specified for the conditions in the gasifier but was gifted to the project. The refractory was mixed into a stiff slurry with water at a rate of 5-7 kg per 100 kg of refractory. The refractory layer in the combustion column is about 50 mm thick giving an internal diameter of 107 mm. The gasification column has a refractory layer about 30 mm thick with an internal diameter of 207 mm. The refractory was vibrated into place on a vibrating table such that water did not separate out from the cement but enough that it flowed into all areas of the cast.

Each of the sections has many ports for thermocouples, pressure tapings, pilot burners and air jets, these are cast into the refractory during construction. Each section contains about six thermocouple ports distributed evenly around the section. The ports all have refractory lining which was moulded during casting of the refractory. The

bottom sections of each column are both quite complex casts containing many ports for chute inlet and the air jets in the combustion column and chute outlet and fuel feed port in the gasification column. Because of these complexities they are cast with Kaowool insulation instead of the vermiculite as it is much easier to work in confined areas and has more structure in small portions.

The refractory attains full strength after drying and firing processes have been undertaken. Initial drying takes place with the cardboard tubes still in place in a dry environment at room temperature. During this part of the process the refractory attains a solid structure and the cardboard cast can be removed. The refractory still contains high levels of moisture which if driven off too quickly will cause the refractory to crack and loose structural integrity. The sections are then heated individually with a gas burner to over 300 °C to drive off water still in the refractory and vermiculite insulation. To help vapour deep inside the insulation escape to the surface when heated, holes were drilled in the steel shell. During the heating with a gas burner the temperature of the cast is monitored so it is not heated too quickly. (Cardoso et al., 2004) show that moisture is driven from the refractory in three phases. The temperature of the cast sections is increased very slowly in these temperature ranges, at rate of 20 °C/h. The final stage of the firing was undertaken once the sections were assembled into columns and were fired with the main LPG burners. The sections are assembled with eight M12 x 60 mm bolts. They are sealed with 10 mm Kaowool blanket gaskets between the refractory to allow for expansion under hot conditions. The sections are sealed gas tight with a high temperature silicon gasket sealant (Ados gasket 260).

Stainless Steel Work

Much of the pipe work and duct work including the cyclones and heat exchangers are constructed from 253 MA grade stainless steel. With an operating temperature of up to 1200 °C 253 MA grade stainless is more than adequate for this application. The drawings for these sections of the gasifier were made by Rick Dobbs and they were constructed and welded by Charlston Engineering. The working drawings can be found in the ‘Gasifier Operational Guide’. The pipe work is constructed from standard BSP pipe in sections so it can be easily taken apart. The fittings and thread is all painted with nickel anti-sieze compound (ROCOL brand).

Drawings

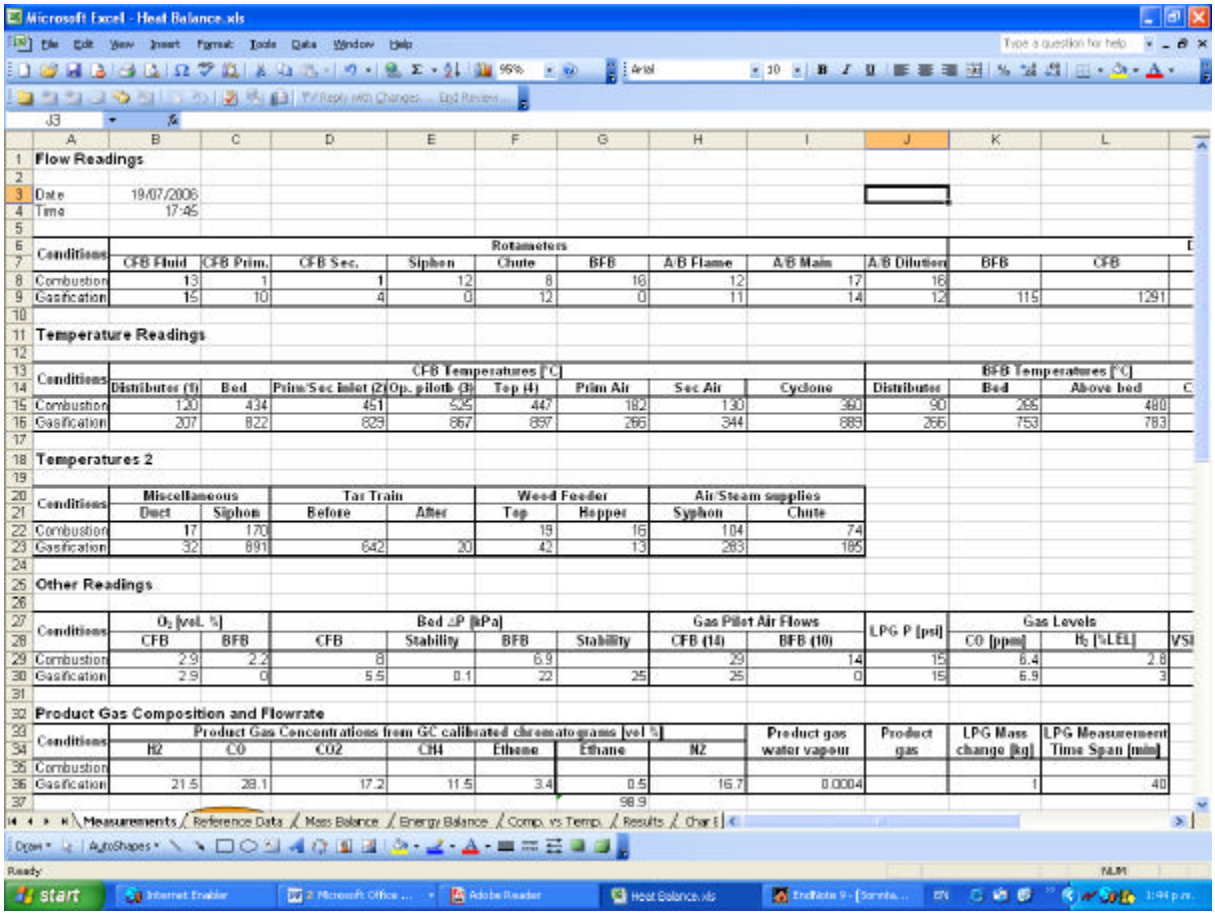
The detailed engineering drawing used to construct the gasifier are available on the gasifier group folder on the chemshared drive of the engineering network.

D. Appendix D

Energy and Mass Balance

Measurement Input

Measurements of flow-rates from rotameters for the air and steam flows and temperatures are recorded manually in data spread sheets. Product gas composition is calculated by the Agilent Cerity QAQC software after calibration. Moisture contents are measured and the mass of water collected is recorded and the concentrations in the product gases are calculate on an ideal gas basis. All the recorded values are input into the data entry "Measurements" worksheet in the heat and mass balance shown below in Figure 0-1.



Microsoft Excel - Heat Balance.xls

Flow Readings

Date	19/07/2006
Time	17:45

Conditions	Rotameters										
	CFB Fluid	CFB Prim.	CFB Sec.	Siphon	Chute	BFB	A/B Flame	A/B Main	A/B Dilution	BFB	CFB
Combustion	13	1	1	12	8	16	12	17	16		
Gasification	15	10	4	0	12	0	11	14	12	115	1291

Temperature Readings

Conditions	CFB Temperatures [°C]								BFB Temperatures [°C]			
	Distributor (1)	Bed	Prim/Sec Inlet (2)	Op. pilot (3)	Top (4)	Prim Air	Sec Air	Cyclone	Distributor	Bed	Above bed	C
Combustion	120	434	451	525	447	182	130	360	90	295	480	
Gasification	207	822	829	867	897	266	344	889	266	753	783	

Temperatures 2

Conditions	Miscellaneous		Tar Train		Wood Feeder		Air/Steam supplies	
	Duct	Siphon	Before	After	Top	Hopper	Syphon	Chute
Combustion	17	170			19	16	104	74
Gasification	32	891	642	20	42	13	283	185

Other Readings

Conditions	O_2 [vol. %]		Bed ΔP [kPa]				Gas Pilot Air Flows		LPG P [psi]	Gas Levels		
	CFB	BFB	CFB	Stability	BFB	Stability	CFB (14)	BFB (10)		CO [ppm]	H_2 [%LEL]	VSI
Combustion	2.9	2.2	8		6.9		29	14	15	8.4	2.8	
Gasification	2.9	0	5.5	0.1	22	25	25	0	15	6.9	3	

Product Gas Composition and Flowrate

Conditions	Product Gas Concentrations from GC calibrated chromatograms [vol %]							Product gas water vapour	Product gas	LPG Mass change [kg]	LPG Measurement Time Span [min]
	H2	CO	CO2	CH4	Ethene	Ethane	N2				
Combustion											
Gasification	21.5	28.1	17.2	11.5	3.4	0.5	16.7	0.0004		1	40
						98.9					

Measurements / Reference Data / Mass Balance / Energy Balance / Comp. vs Temp. / Results / Chart

Ready

Figure 0-1. Data entry sheet of heat and mass balance.

i. Heat Capacities, Heat of Formations and Properties

Heat capacity for a particular species varies with temperature. Therefore, in order to accurately calculate the enthalpy changes in the system, this dependence of heat capacity on temperature needs to be recognized. The literature suggests using the following correlation for heat capacities:

$$CP_i = R\left[A_i + B_i \frac{T_{reaction} + T_{amb}}{2} + \frac{C_i}{3}\left(T_{reaction}{}^2 + T_{amb}T_{reaction} + T_{amb}{}^2\right) + \frac{D_i}{T_{reaction}T_{amb}} \right]$$

Sonntag et.al (1998)

Where the constants Ai, Bi, Ci and Di can be found in thermodynamic tables and are given below:

Table 0-1. Heat Capacity Correlation Constants

	Heat Capacity Constants					Heat of Formation
	Tmax (K)	**A**	**B**	**C**	**D**	**Ho (kJ/kmol)**
Methane	1500	1.70	9.08E-03	-2.16E-06	0.00E+00	-74520.0
Hydrogen	3000	3.25	4.22E-04	0.00E+00	8.30E+03	0.0
Carbon Monoxide	2500	3.38	5.57E-04	0.00E+00	-3.10E+03	-110525.0
Carbon Dioxide	2000	5.46	1.05E-03	0.00E+00	-1.16E+05	-393509.0
Water Vapour	2000	3.47	1.45E-03	0.00E+00	1.21E+04	-241818.0
Water (liquid)						-285830.0
Ethene	1500	1.13	1.92E-02	-5.56E-06	0.00E+00	52510.0
Ethane	1500	1.42	1.44E-02	-4.39E-06	0.00E+00	-83820.0
Carbon, Char	2000	1.77	7.71E-04	0.00E+00	-8.67E+04	0.0
Nitrogen	2000	3.28	5.93E-04	0.00E+00	4.00E+03	0.0
Oxygen	2000	3.64	5.06E-04	0.00E+00	-2.27E+04	0.0

These constants are valid up to at least 1500 K and were extracted from Smith, Van Ness and Abbott (2001). Heat capacity of sand was extracted from Sonntag et al. (1998) and assumed to be constant with temperature.

Heat Capacity of Sand = 0.8 kJ/kg/K.

The heat and mass balance uses look up tables for the properties of the gases. Tables for the properties of air were extracted from Mills (1999). Steam tables used in the heat balance were extracted from Tucker (1999).

ii. Mass and Energy Balance Calculations

The energy balance uses the product gas composition to find the energy transferred in the bed material from the combustion column to the gasification column. The heat balance was primarily set up to calculate the bed material circulation rate under gasification conditions; however, it is useful for calculating the char circulation rate and getting the data into a useful format. The balances uses the product gas composition (from the micro GC) and the moisture content of the product gases to calculate a heat of formation and sensible heat required to form the gas. The heat required for the reactions is assumed to be all supplied to the gasification column in hot bed material circulating from the combustion column. The energy balance uses solver to find the bed material circulation rate supply the required heat to generate the product gas. The char circulation rate is calculated uses a mass balance on the combustion column. The oxygen concentration and LPG feed rate are recorded and oxygen not combusted by the LPG is assumed to be combusted by char circulated from the gasification column. The balance works by using solver to find the char circulated to balance the oxygen. Solver is set and runs twice to initially calculate the char circulation rate then calculate the bed material circulation rate. The product gas flow-rate is calculated with an elemental balance on the carbon in the system required for the energy balance used to find the bed material circulation rate. Figure Figure 0-2 shows the energy balance used to calculate the bed material circulation rate. Figure Figure 0-3 shows the mass balance spreadsheet used to calculate the char circulation rate and the product gas flow rate.

Inputs to BFB

Steam

Heat of formation of steam	-241818	kJ/kmol
Proportion of syphon steam to exhaust	0.1	
Proportion of chute steam into CFB	0.4	

	Flowrate [mL/min]	Water Temperature [°C]	Mass flow [kg/min]	Mass Flow to BFB [kg/min]	Steam Temperature [°C]	Cp [kJ/kg.K]	Sensible Heat [kJ/min]	Molar flow [kmol/min]	Heat of Formation
Fluidising	170	25	0.169626	0.169626	266	1.981	80.983015	0.00942367	-2278.81
Syphon	0	25	0	0	283	2.001	0	0	0
Chute	6	25	0.0059868	0.00239472	185	1.975	0.7567315	0.00013304	-32.1715
Total			0.1756128	0.17202072			81.739746		-2310.98

Wood

Heat of formation of wood	-126782.44	kJ/kmol
Molecular Mass	24.02	kg/kmol

Jacks spread sheet for pellets of composition $CH_{1.44}O_{0.68}$

Assume wood is fed at the reference temperature so it does not have any sensible heat effects

Assume feed stock is pellets so minimal moisture content. if chips are fed this may have to be taken into account.

Heat of Formation

	Auger Setting	Mass Flow [kg/min]	Mass Flow [kg/s]	Molar Flow [kmol/min]	ΔHf wood [kJ/min]
Wood Auger	1.6	0.2508	0.00418	0.0104	-1323.77
Wood Moisture		0.0201		0.0011	

Nitrogen Purge

No heat of formation only contribution from sensible heat

	Flow rate [L/min]	Temperature [°C]	Mass flow [kg/min]	Molar flow [kmol/min]	Cp [kJ/kmol.K]	Sensible heat [kJ/min]
Nitrogen	0.3	13	0.054	0.003	-1088.05	43.61

Outputs from BFB

Product gas flowrate	15.615978 L/min

Use Ideal gas law to find the molar flow rate. Check this using some other model such as HYSYS.

PV=nRT

Pressure	110 kPa
Molar flowrate	0.00020137 kmol/min

Product gas

	Molecular Weight [kg/kmol]	Composition dry basis [vol. %]	Composition wet basis [vol. %]	Flow rate [L/min]	Temperature [°C]	Molar flow [kmol/min]	Molar flow [kmol/s]	Mass flow [kg/min]	Cp [kJ/kmol.K]		Sensible heat [kJ/min]	Heat of Formation [kJ/min]
Hydrogen	2	21.5	14.2	2.2242	753	0.0025	0.000042	0.01	30.68	4.37	56.4	0.0
Methane	16	11.5	7.6	1.1897	753	0.0014	0.000023	0.02	72.67	10.36	73.9	-100.7
Carbon Dioxide	44	17.2	11.4	1.7794	753	0.0020	0.000034	0.09	53.39	7.60	80.3	-795.0
Carbon Monoxide	28	28.1	18.6	2.9070	753	0.0033	0.000055	0.09	32.79	4.67	81.5	-364.8
Ethene	28	3.4	2.3	0.3517	753	0.0004	0.000007	0.01	124.73	17.77	37.5	21.0
Ethane	30	0.5	0.3	0.0517	753	0.0001	0.000001	0.00	96.18	13.70	4.3	-4.9
Nitrogen	28	16.7	11.1	1.7276	753	0.0020	0.000033	0.05	32.36	4.61	47.8	0.0
Steam	18		33.8	5.2708	753	0.0060	0.000100	0.11	41.31	5.88	186.2	-1447.1
								23.0125691		68.95	567.8	-2691.4

	Mass [kg]	Temperature of sample [°C]	Volume of sample [L]	Volume of steam [L]	Ratio steam/total
Water vapour	0.0004	642	5	1.69E+00	0.338

Char

Assume BFB bed is perfectly mixed so temperature of solids leaving bed are at the same temperature as rest of bed.

	Molar flowrate [kmol/hr]	Temperature [°C]	Cp [kJ/kmol.K]	Sensible heat [kJ/min]
Char	0.0962	753	20.62	24.07

Sand Circulation Rate

$$\dot{m}_{sand} = \frac{\dot{m}_{char} C_{p,char} \left(T_{in} - T_{ref}\right)_{char} + H_{f,BFBgas} + \dot{m}_{BFBgas} C_{p,BFBgas} \left(T_{in} - T_{ref}\right)_{BFBgas} - H_{f,wood} - \dot{m}_{steam} C_{p,steam} \left(T_{in} - T_{ref}\right)_{steam}}{C_{Psand} \left(T_{in} - T_{out}\right)_{sand}}$$

125					
126	Char sensible heat	24.1	kJ/min	0.401	kW
127	BFB gas heat of formation	-2691.4	kJ/min	-44.857	
128	BFB gas sensible heat	567.8	kJ/min	9.464	
129	Wood heat of formation	-1323.8	kJ/min	-22.063	
130	Steam heat of formation	-2311.0	kJ/min	-38.516	
131	Steam sensible heat	81.7	kJ/min	1.362	
132	Total	1453.5	kJ/min	24.225	kW
133					
134	Heat Capacity [J/kg.K]	800	J/kg.K. This is an approximate value from Ian, will need a more credible reference		
135	Temperature in (Syphon)	891	°C		
136	Temperature out (BFB bed) [K]	753	°C	24.22	kW heat circulated by bed material
137					
138	Mass Circulation rate of sand	0.22	kg/s		
139		790	kg/hr		
140					

Figure 0-2. Energy Balance Spreadsheet

Mass Balance Under Gasification Conditions

Mass Balance for CFB under Gasification conditions

Inputs
Outputs
Parameters

Run Solver and mass balance

Ratio of Propane to Butane

			vol. fr.	M.W. kg/kmol
Propane	Alpha	α	1	42
Butane	Beta	β	0	58

Measurements

Air Flows

Molecular weight of air: 28.6 kg/kmol =VLOOKUP(A69+273,dryair,9)+((VLOOKUP((50+A69+273),dryair,9)-VLOOKUP((A69+273),dryair,9))/50)*(A

	Rotameter Type	Rotameter Reading [cm]	Actual Flow [m³/s]	Mass Flow @ 40 °C [kg/h]	Temperature [°C]	Density [kg/m³]	Mass Flow [kg/s]	Mass Flow [kg/h]	Molar flow [kmol/h]	O2 in [kmol/hr]
Fluidising air	24K	15	0.00246	2.76E-03	207	7.31E-01	1.80E-03	9.92E+00	3.47E-01	7.25E-02
Primary air	18K	10	0.00119	1.33E-03	266	6.47E-01	7.71E-04	4.80E+00	1.68E-01	3.51E-02
Secondary air	47K	4	0.00636	7.11E-03	344	5.65E-01	3.60E-03	2.56E+01	8.96E-01	1.87E-01
Pilot Burner Air	14K	25	0.00144	1.60E-03	0	1.29E+00	1.85E-03	6.78E+00	2.02E-01	4.22E-02
Total Air								46.105	1.612	0.337

Syphon/Chute Steam Flows

Molecular weight of steam: 18 kg/kmol

Under gasification conditions the syphon and chute are fed with steam or Air some of which is carried into the CFB

	Cold Water Flow [mL/min]	Water Temperature [°C]	Density [kg/m³]	Density [kg/mL]	Mass Flow [kg/h]	Molar flow [kmol/h]
Syphon Steam	25	20	998.0	0.000998004	1.49700599	0.083167
Chute Steam	6	20	998.0	0.000998004	0.35928144	0.01996008
Total Steam					1.85628743	0.10312709

Proportion of Syphon and Chute fluidising agent into CFB: 0.5
Proportion of Syphon and Chute fluidising agent into CFB: 0.5

Steam flow to combustion side of gasifier: 9.2814E-01 0.0000E+00

This is equal to the proportion multiplied by the total steam flow as for air

Syhon/Chute Air

	Rotameter Type	Rotameter Reading [cm]	Actual Flow [m³/s]	Mass flow @ 40 °C [kg/h]	Temperature [°C]	Density [kg/m³]	Mass Flow [kg/s]	Mass Flow [kg/h]	Molar flow [kmol/h]	Vol/hr
Syphon Air	14K	0	0.00017	1.95E-04	0	1.29E+00	2.24E-04	7.00E-01	2.45E-02	
Chute Air	14K	12	0.00070	7.83E-04	0	1.29E+00	9.02E-04	2.82E+00	9.86E-02	
Molar Proportion of Syphon and Chute air combust in BFB									0.7	
Total air								2.464	0.686	

Percentage O2

Inlet air: 20.6 vol.%
Combustion flue gases: 2.9 vol.%
0.029 mol fr

LPG

Molecular Weight	42	kg/kmol
Mass change	1	kg
Rotameter	8	min
Measured Vol. flow	9.84	L/min
	0.029	kg/min
	1.75	kg/h
Molar Flow	0.0417	kmol/h

Char from BFB (gasification reactions)

Molar flow	0.096	kmol/h
Molecular Weight	12	kg/kmol
Mass flow	1.154	kg/h

Elemental Balance

to estimate the composition of the combustion gases and hence calculate the amount of char combustion

Assumptions

Air composition

| | | |

	VF Drive	Wood Flow	Wood Flow	Component weight %					Component Flows [kg/h]				
	Setting	rate [kg/s]	rate [kg/h]	C	O	H	N	ash	C	O	H	N	ash
	1.6	0.00418	15.048	51.3	42.4	5.81	0.08	0.4	7.719624	6.380352	0.8742888	0.0120384	0.060192
									Component Flows [kmol/h]				
									C	O	H	N	ash
									0.643302	0.398772	0.8742888	0.0008599	0.0015048

Biomass / Steam Ratio

Biomass	15.05	kg/h	
Steam	12.04	kg/h	
Ratio	1.25	kg/kg	

Nitrogen Purge

	Vol. Flow	Density @ 300 K	Mass Flow		Molar Flow
	[L/min]	[kg/m³]	[kg/min]	[kg/h]	[kmol/h]
Nitrogen	12	1.16	0.01392	0.8352	0.029828571

Product Gas

Duct dimensions

Height	0.07	m
Width	0.18	m
CS Area	0.0126	m²

Dry Product Gas GC Component Analysis [mol %]					Product gas	Product gas	Dry Product Gas GC Component Flow [m³/h]				
H2	CO	CO2	CH4	N2	velocity [m/s]	flow [m³/s]	H2	CO	CO2	CH4	N2
21.5	28.1	17.2	11.6	16.7	1.239	0.016	0.336	0.439	0.269	0.180	0.261

Elemental Balance

to estimate the composition of the combustion gases and hence calculate the amount of char combustion

Assumptions

Air composition

O2	21	%
N2	79	%

Complete combustion of the LPG and char to H2O and CO2

ideal gas so mol% = vol% from O2 sensor

	Steam	Air	Wood	Prod. Gas	Comb. Out		
C	0.000	0.000	0.643	0.000			
H	0.000	0.000	0.874	0.000			
O	0.000	0.522	0.399	0.522			
N	0.000	0.522	0.001	0.522			
Ash	0.000	0.000	0.002	0.000			
H2	0.000	0.000	0.000	0.336			
CH4	0.000	0.000	0.000	0.180			
CO2	0.000	0.000	0.000	0.269			
CO	0.000	0.000	0.000	0.439	0.439		
O2	0.000	0.069	0.000	0.000	0.000		
N2	0.000	0.261	0.000	0.920	0.920		
H2O	0.000	0.000	0.017	0.261	0.261		
				total mol	1.620	mol fr O2	0.271

Char will include some steam from the fluidising steam in the chute

Carbon Balance to Calculate Product gas flow rate

Assume all carbon into the BFB come in the form of wood

So no carry over of combustion gases or LPG from the CFB

					Percentage C in biomas / C circulating
Molar flow of C into gasifier	0.643	kmol/h	7.719624	kg/h	0.1496
Char flow out of BFB	0.096	kmol/h	1.154	kg/h	
Carbon in Product Gases	0.547	kmol/h	6.565	kg/h	

Carbon out of BFB			No C atoms	Amount	
Methane	7.6	vol%	1	7.6	0.097
Carbon Dioxide	11.4	vol%	1	11.4	0.146
Carbon Monoxide	18.6	vol%	1	18.6	0.238
Ethene	2.3	vol%	2	4.5	0.058
Ethane	0.3	vol%	2	0.7	0.008
Total	40.2	vol%		42.8	0.547
			Ratio	0.0128	

426	Carbon flows in product gas on a dry basis			
427	Methane	0.0974	kmol/h	
428	Carbon Dioxide	0.1457	kmol/h	
429	Carbon Monoxide	0.2380	kmol/h	
430	Ethene	0.0288	kmol/h	
431	Ethane	0.0042	kmol/h	
432	Nitrogen	0.2749	kmol/h	
433	Hydrogen	0.2749	kmol/h	
434	Total	1.0638	kmol/h	
435				
436	Use Ideal Gas equation to find the volumetric flow rate			
437	V=nRT/P			
438				
439	Temperature of gas (BFB Cyclone)	642	°C	
440	Gas Constant, R	8.314		
441	Pressure, Atmospheric	101	kPa	
442				
443	Volume flow	56.22	m³/h	
444				
445	Velocity	1.239	m/s	
446				
447	**Air Flow into BFB with Wood Feed**			
448				
449	Use a nitrogen balance with GC analysis and know purge flow rate to calculate air flow into BFB			
450	Assume no Nitrogen reacted or leaked			
451				
452		Molar Flow [kmol/h]	Mass Flow [kg/h]	
453	Nitrogen Product gas	0.27	7.70	
454	Nitrogen in Purge	0.03	0.8352	
455	Nitrogen in Air to Syphon and Chute	0.09	2.464	
456	Nitrogen from Air fed with wood	0.16	4.40	
457	**Air fed with Wood**	0.20	5.83	
458				
459				

Figure 0-3. Mass Balance Spreadsheet.

iii. Results

The results are printed into the "Results" worksheet by the same VBA program which runs solver to calculate the char and bed material circulation rates.

	A	B	C	D	E	F	G	H	I	J	K	L	M	N	O	P
1				Bubbling Fluidised Bed												
2				Conditions of Operation												
3				Inputs								Operating Temperatures [°C]				Char
4	Date	Time	Gasification or Combustio	Wood Feed [kg/h]	LPG Feed rate [kg/h]	Total Steam [kg/h]	Steam/Biomass Ratio [kg/kg]	Nitrogen Purge [kg/h]	Air in wood / chute / syphon	Bed Material Circulation Rate [kg/h]	Heat Transfer into Bed [kW]	Bed	Above Bed	Steam	Product Gas @ Cyclone	Circulation Rate [kg/h]
5	19-Jul-08	3.00.00 p.m.	Gasification	15.05	0.00	10.10	0.68	0.70	9.40	613	19.22	750	770	300	472	191
6	19-Jul-08	3.30.00 p.m.	Gasification	15.05	0.00	10.10	0.68	0.70	7.60	580	18.18	760	776	300	472	224
7	19-Jul-08	4.00.00 p.m.	Gasification	15.05	0.00	7.80	0.48	0.68	4.31	586	6.00	693	726	190	580	347
8	19-Jul-08	4.45.00 p.m.	Gasification	15.05	0.00	7.80	0.48	0.68	6.19	124	2.98	721	735	170	585	352
9	19-Jul-08	5.00.00 p.m.	Gasification	15.05	0.00	9.50	0.64	0.68	3.91	535	15.89	694	720	220	600	254
10	19-Jul-08	5.30.00 p.m.	Gasification	15.05	0.00	9.50	0.64	0.68	5.60	567	28.03	686	730	250	610	176
11	19-Jul-08	5.50.00 p.m.	Gasification	15.05	0.00	10.54	0.70	0.68	5.71	779	27.52	734	781	260	626	166
12	19-Jul-08	6.45.00 p.m.	Gasification	15.05	0.00	12.04	0.80	0.68	5.60	790	24.22	760	790	280	642	116
13																
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Figure 0-4. Results spreadsheet shows all data collects and circulation rates calculated

E. Appendix E

CAPE Gasifier HAZOP Safety Review

HAZOP Meeting Minutes

The following is the minutes recorded for the Univeristy of Canterbury, Department of Chemical and Process Engineering, FICFB gasifier HAZOP meeting on 20 and 21 December 2005. This was a major part of the construction phase of the gasification plant.

Job No.:			Meeting No:	
Job Title:	CAPE Gasifier HAZOP/Safety Review			
Meeting held at:	University of Canterbury		Date of Meeting:	20-21 Dec. 2005
Subject:	HAZOP			
Companies:	*Delta S Technologies and University of Canterbury*			
Present:	See Minute Items 2 and 9			
Recorded by:	Eric Scharpf	Signed:	Date:	21 December 2005
Accepted by:		Signed:	Date:	
Distribution:	Preliminary version (Rev P) for team review and comment			

1.	General Introduction Eric Scharpf	Date to be completed	Person Responsible
2.	20 December Participant Introductions: Rick Dobbs – Design Engineer, Supervised Construction, Lead Instrumentation/Project Engineer Shusheng Pang – Canterbury Lead Researcher/Process Eng /Project Manager Jack Rutherford – Canterbury Tech Scribe/Operations Eric Scharpf– Delta S Tech: Safety lifecycle facilitator. Chris Williamson – Process Engineer (Part-time) Ian Gilmour – Process Engineer (Leader of Gas Cleaning Objective) Bob Gordon – Senior Technical Officer, Electrical safety Frank Weerts – Mechanical Technician Peter Gostomski – HOD Cape (Observer) Trevor Berry – Technician, Deputy Safety Officer John Abrahamson – Chairman of CAPE UoC Safety Committee, cyclone design consultant. David Brown – CAPE Safety Officer (UoC) Jock Brown – Operator		

3.	Eric Scharpf introduced the method and background for the workshop.		
4.	HAZOP/Safety Review will consider – CAPE Two Stage Fluidised Bed Wood Combustor as recorded on the P&ID document Flowdiagram2.pdf (undated) attached as part of this report with HAZOP nodes marked.		
5.	Documentation used in the Review- 1. Piping and Inst. Diagram (Flowdiagram2.pdf undated) 2. Risk Analysis Update (undated) 3. Safety Briefing Sheet (undated) 4. Training Slide Handouts (14 Dec 2005) 5. HAZOP worksheet example (14 Dec 2005)		
6.	The workshop will consider the gasifier in its current condition of preliminary operation as of 20 December 2005. It will identify hazards along with existing and proposed safeguard layers of protection and safety instrumented functions. Since it will only qualitatively address accident likelihood and since tolerable risk has not been rigorously defined, the review will not attempt to make any determination of whether an additional safety instrumented function is needed or what SIL may be required.		
7.	The group used the following 8 parameters to analyse the gasifier: Flow (Air, hot gas, water/steam, LPG, wood, sand) Temperature Power and Instrument Air Failure Other parameters were determined to be redundant with the guide words or not relevant to the application of the gasifier.		
8.	On the 20th, the group covered the first node (afterburner and heat exchangers 4-6) and made a start on the second node (CFB and exchangers 1-3) flow parameter before leaving for the day.		
9.	21 December Participant Introductions: Rick Dobbs – Design Engineer, Supervised Construction, Lead Instrumentation/Project Engineer Shusheng Pang – Canterbury Lead Researcher/Process Eng /Project Manager Jack Rutherford – Canterbury Tech Scribe/Operations Eric Scharpf– Delta S Tech: Safety lifecycle facilitator.		

	Chris Williamson – Process Engineer (Part-time) Ian Gilmour – Process Engineer (Leader of Gas Cleaning Objective) Bob Gordon – Senior Technical Officer, Electrical safety Trevor Berry – Technician, Deputy Safety Officer (Afternoon) Frank Weerts – Mechanical Technician John Abrahamson – Chairman of CAPE UoC Safety Committee, cyclone design consultant. David Brown – CAPE Safety Officer (UoC) Jock Brown – Operator		
10.	On the 21st, the group reviewed the first node and then worked through the remaining parts of the P&ID with a final review of work to assign responsibility for each of the actions items identified from the workshop.		
11.	The flame supervision system is critical to the overall safety of this system. The effectiveness of this system is unknown and should be evaluated.		(SP, RD, ES & IAG)
12.	Consider emergency lighting for power failure.		BG & DB
13.	Controller is open access and potential for unauthorised use/error exist. Consider proper procedure protection.		Health and Safety Committee
14.	If flame supervision is altered to enable operation without pilots, further evaluation will be needed. With particular reference to low (dropping) temperatures in CFB.		(Should be considered in conjunction with item 11)
15.	Consideration of heat sensitive paint for indication of long term refractory breakdown.		(IAG)
16.	The Electric Arc Reactor was not a part of this study. Safety protocols for the unit exist independent of this report.		
17.	Inert gas and H2/CO/CH4 mix handling procedures are not a part of this study and should be investigated separately.		(JWB, JPR & TB)

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Unit:	CAPE						
ITEM Number	**PARA METER**	**GUIDE WORD**	**CAUSES**	**CONSEQUENCES**	**PROPOSED (P) & EXISTING (E) SAFEGUARDS MSF/SIF/Other**	**SENSOR/ FINAL ELEMENT Tag**	**COMMENTS ACTIONS**
	Flow (water)	***NONE***	Blocked line or loss of feed pressure. Pressure relief vented. Build-up of scale in HX pipes.	Overheat of heat exchangers and hot gas released to extraction duct. Note: Exchanger duty is small relative to temperature of gas. Potential baghouse fire. See other hazards.	(E1O) Operator action on high temperature/low pressure drop across BFB to add water or shut off system (E2MSF) system designed to handle hot gas (E3MSF) Duct fan carries away hot gas. (E4SIF) Hot temperature cut out in duct.		System may change in relation to proposal to automate pressure drop detection in BFB. **Investigate potential of fire in baghouse further** (To be done in association with Fire Eng – **IAG and DB**). See other hazards.

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		MORE	Valves fail open. Change in delivery pressure of water. Low pressure water delivery system changed over to high pressure system.	Liquid water fed to BFB. Minor safety issues.	(E1O) Operator action on low temperature or high flow or high pressure across bed (E2MSF) Pressure relief should pressure drop across BFB rise. (E3O) Procedure for changeover		
		LESS	See NONE				
		PARTLY	See NONE				
		AS WELL AS (Air)	Procedural error	See BFB node			

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	Flow (water)	***REVERSE***	Pressure relief vessel is empty	Hot gases (possibly LPG during start-up) vented through pressure relief into room.	(E1MSF) Extraction fan can cope with expected hot gas (&/or LPG) release. (E2O) Lack of pressure drop across BFB and low oxygen in BFB and dropping temperature in BFB. (E3SIF) LPG cuts out on low oxygen Note initiating event assumes failure of relief valve protection layer (E4SIF) Existing CO and H2 monitoring which cuts all fuels (wood and LPG)		**Consider adding LPG detection in the room (BG).** **Investigate possibility of automatic refill of pressure relief (JWB & FW)**.

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		OTHER THAN	Extremely unlikely				
		PART OF	See NONE				
		CHANGE	Poor preliminary check of line connections.	Water leak or hot gas leak into room. See above as well.	(E1O) Rigorous cold check procedure. (E2O) Numerous instruments show leaks during start-up on air. Note: Interlocks may or may not prevent accident		**Consider additional safeguards (See LPG detection)**

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	Flow (hot gas)	***NONE***	Rootes blower failed causes total air loss. Valve failure causing loss of air to A/B only..	Combustible gas into duct system	(E1SIF) Flame failure cuts off A/B and BFB LPG. (E2SIF) Oxygen measured on afterburner and automatically cuts wood to BFB and LPG. (E3O) Numerous instruments led to operator shut down of plant. (E4SIF) Temperature controlled and cuts out at high temp (Uncertain in this case) (E5SIF) Flow meter on total air flow to a/b. Low flow cutout. Special Note: SIF share common final elements.		**Check speed of E1SIF (Burnout period of both char and gases) (JWB)** **Check concentration of gases versus LEL in the duct (JWB and JPR)** **Consider feedback signal from Rootes Blower (RD & BG)**

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	Flow (hot gas)	***MORE***		Potential loss of pilot flame. Unburned gas situation unlikely. Potential baghouse fire from sparks and/or char blow through.	(E1SIF) Delay on restart of pilot until BFB has burnt out. (E2MSF) Cyclones prevent particle carry over (E3MSF) Sand trap prevents particle carry over (E4MSF) Long residence time in duct allows particles to cool.		**Investigate potential of fire in baghouse further (See IAG)**. Discuss with Fire Engineering.
		LESS	Valve in particle laboratory opened more than 1/3	See NONE			
		PARTLY		See NONE			

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		AS WELL AS	Not likely				

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	Flow (hot gas)	***REVERSE***	Failure of Rootes Blower while valve is open in the particle laboratory	Possible hot combustible and toxic gas into particle laboratory	(E1SIF) Flame failure cuts A/B & BFB LPG and wood (E2SIF) Oxygen measured on AB and automatically cuts LPG and wood. (E3O) Numerous instruments led to operator shut down of plant. (E4SIF) Temperature controlled and cuts out at high temp (Uncertain in this case) (E5SIF) Flow meter on total air flow to a/b. Low flow cutout. Special Note: SIF share common final elements. **(P6O) Run exhaust fans in**		**Consider adding check valve on main air line (JWB and FW)**. Most existing safeguards do not eliminate the hazard. **Investigate this situation in more detail off-line if check valve is not installed (JWB)** **Check speed of E1SIF (Burnout period of both char and gases) (JWB and JPR)** **Add running exhaust fans (P6O) to operating procedure (JWB)**

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		PART OF	See None				

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	Flow (hot gas)	***CHANGE***	Bad seal of A/B distributor, air lines.	Leakage of LPG or air (air is cold during start-up and becomes hot during operation) into room	(E1O) Flame failure will pick up if there is no ignition. This is not all inclusive. (E2O) Operator may smell LPG or may feel air leak during start-up. (E3MSF) Large air flow through exhaust system (E4MSF) Two extraction fans from the room (E5MSF) Two convection vents in the room. **(P6O) Procedure to leak test whenever LPG system is opened.**		Investigate this further. Consider adding LPG sensor in room. (See elsewhere) **Health and Safety committee to consider P6O to general LPG policy.**

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	Flow of LPG	***NONE***	Valve failure. Cylinders empty.	This can result in loss of A/B flame in afterburner. Can send producer gas into duct. LPG could come back on. Failure of the flame supervision will send LPG to the duct.	(E1MSF) Producer gas will continue to burn in afterburner. (E2SIF) Loss of pilot cuts out all fuel and locks out LPG for 2mins. (E3O) Temperature indication in the A/B		**Investigate flame limits in the A/B and duct (JWB and JPR)**. **Check burnout time of producer gas.** (See elsewhere)

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	Flow of LPG	***MORE***	Small reducer failure.	Higher temperature. Surge may shut out pilot and build up unburned gas.	(E1SIF) LPG cut off on loss of pilot. (E2SIF) All fuels cut out on high temperature in A/B (E3O) Numerous indications to operator of fault		Reconsider if other users of LPG come on line.
		LESS	See NONE				
		PARTLY	See NONE				
		AS WELL AS	Not Relevant				
		REVERSE	Not relevant				
		OTHER THAN	See Change				
		PART OF	See NONE				

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		CHANGE	Bad connection of LPG lines or change in fuel grade.	Leakage of LPG into room	(E1O) Operator may smell it or may not. (E2MSF) Large air flow through exhaust system (E3MSF) Two extraction fans from the room (E4MSF) Two convection vents in the room. **(P5O) Procedure to leak test whenever LPG system is opened.** (See elsewhere)		Investigate this further. Consider adding LPG sensor in room (See elsewhere). **Check if it is possible to attach acetylene (DB)**.
	Temp. (water/hot gas)	***NONE***	Not relevant				

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		MORE	Wrong air/fuel balance	Equipment damage due to high temp. Refractory KaO wool damage. Minimal safety risk	(E1SIF) Automatic temperature cut-out and alarm (E2O) Operator detect numerous indications		
		LESS		Unburned gas see FLOW.			
		PARTLY	Not relevant				
		AS WELL AS	Not relevant				
		REVERSE	Not relevant				
		OTHER THAN	Not relevant				
		PART OF	Not relevant				

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		CHANGE	Not relevant				

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	Power and/or Instrument Air Failure			Combustible gas will build up in the room as extraction system is lost.	(E1MSF) Two convection vents exist in room (E2SIF) Controller locks out to prevent automatic restart. **(P3O) Operator procedure for power failure**		Investigate amount of combustible and toxic material released to laboratory (see burnout comment). **Check behavior of services on power return (BG)**. **Instrument air system/Compressed air system under review (BG and FW)**. **Consider Potential interlock (RD).** **Consider emergency lighting (BG and DB)**. **Operator procedure for power failure (JWB &**

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NODE Number	CFB Air HEX 1-3	**DRAWING NUMBER:**	P & I Diagram CAPE Gasifier	**NODE DESCRIPTION:**	Provides the thermal energy for the process through the combustion of char and LPG.		
Unit:	CAPE						
ITEM Number	**PARA METER**	**GUIDE WORD**	**CAUSES**	**CONSEQUENCES**	**PROPOSED (P) & EXISTING (E) SAFEGUARDS MSF/SIF/Other**	**SENSOR/FINAL ELEMENT Tag**	**COMMENTS ACTIONS**
	FLOW (Air through HX 1-3 and CFB)	***NONE***	Rootes Blower fails, valve shut,	Unburned fuel accumulation in CFB and combustible gases into extraction duct. Not as serious as the A/B. Loss of sand circulation.	(E1SIF) Oxygen sensor will cut LPG (E2O) Rotameters and pressure drop will indicate to operator (E3SIF) Automatic cut-out on low/no air flow.		**Response times of oxygen cut-out is 10s max. Undertake explosive quantity calculation (JWB and CW)**. Note: Copy this result to other A/B loss of air.

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		MORE	Valves closed in other rooms or Rootes blower speed altered.	Cyclone blocked and sand, ash and char blown out the HX. Hot particles in room and potential fire hazard.	**(P1MSF) Sand catcher under construction.** (E2O) Rotameters and pressure drop indicate to operator **(P3O) Procedure with Rootes blower**		**(P1MSF) Sand catcher under construction (JWB and FW)** **(P3O) Procedure with Rootes blower (JWB)**

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	FLOW (Air through HX 1-3 and CFB)	***LESS***	Sand can block CFB distributor, ports, cyclone and HX. Air valves opened in other part of building.	See None	(E1SIF) Oxygen sensor will cut LPG (E2O) Rotameters and pressure drop will indicate to operator (E3SIF) Automatic cut-out on low/no air flow. **(P4O) Procedure for only one operator of Rootes blower at any one time** (See elsewhere)**.**		Note: Other building could have future Rootes blower users.
		PARTLY	See Less				

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		AS WELL AS	Not relevant				
		REVERSE	See A/B REVERSE				
		OTHER THAN	Not relevant				
		PART OF	Not relevant				
		CHANGE	Poor air connections when removing distributor (See Less).	Hot air leak	**(P1O) Operating procedure .**		**Formal sign off on operating procedure to be considered for each gasifier start (Health and Safety - JA).**

Design Intent:	To generate medium calorific value fuel gas from biomass in a staged reaction of steam moderated pyrolysis in a bubbling fluidized bed (BFB) and combustion of residual char in a circulating fluidized bed (CFB). The system uses flow of an inert solid (sand) to transfer heat between the two reaction stages. The fuel gas is then combusted in a downstream afterburner.						
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Unit:	CAPE						
ITEM Number	**PARA METER**	**GUIDE WORD**	**CAUSES**	**CONSEQUENCES**	**PROPOSED (P) & EXISTING (E) SAFEGUARDS MSF/SIF/Other**	**SENSOR/FINAL ELEMENT Tag**	**COMMENTS ACTIONS**
	FLOW (Hot gas)	***NONE***	Rootes Blower fails, valve shut,	No safety issue.	(E1SIF) Oxygen sensor will cut LPG (E2O) Rotameters and pressure drop will indicate to operator (E3SIF) Automatic cut-out on low/no air flow.		

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		MORE	Valves closed in other rooms or Rootes blower speed altered.	Cyclone blocked and sand, ash and char blown out the HX. Hot particles in room and potential fire hazard. See air flow.	**(P1MSF) Sand catcher under construction (**See else where)**.** (E2O) Rotameters and pressure drop indicate to operator **(P3O) Procedure with Rootes blower** (See else where)		

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	FLOW (Hot gas)	***LESS***		No safety issue	(E1SIF) Oxygen sensor will cut LPG (E2O) Rotameters and pressure drop will indicate to operator (E3SIF) Automatic cut-out on low/no air flow. **(P4O) Procedure for only one operator of Rootes blower at any one time** (See else where) **.**		Note: Other building could have future Rootes blower users.
		PARTLY	See Less	No safety issue			

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		AS WELL AS 1	Siphon empty allowing flow from CFB to BFB	Air into BFB. Combustion occurring in the top of BFB and in BFB to A/B cyclone. Higher temperatures through this zone and possible meltdown.	(E1O) Operation has some indication of temperature in this zone (but not much). (E2MSF) Flow tends to go through siphon from BFB when siphon is empty. **(P2O) Gas composition slipstream to be taken in this zone (JWB).**		**Investigate adding a temperature probe in the cyclone prior to the A/B (JWB and RD)**. **Investigate pressure drop across siphon (JWB and RD)**.

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	FLOW (Hot gas)	***AS WELL AS 2***	Siphon blocked	Increased carry over of particles into sand catcher. Possible overflow into room or flow restriction. Temperature in BFB will drop.	(E1O) Pressure drop in beds and siphon temperature (and temperature drop in BFB) will indicate fault to operator (E2O) Manual check of siphon at start-up (E3SIF) BFB fuel stopped on low BFB temp		

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		REVERSE	Siphon empty allowing flow from BFB to cyclone and HX	Increased carry over of particles into sand catcher. Possible overflow into room or flow restriction. Some combustion in CFB cyclone and combustible gases released into ducting.	(E1O) Pressure drop in beds and temperatures will indicate fault to operator (E2SIF) CFB oxygen will go to 0 will cut main LPG (E3MSF) Ducting and extraction system designed to handle this size flow.		**Possibility of automating pressure drop in beds control (E1O) (JWB and RD)**. Investigate pressure drop across siphon (See elsewhere)
		OTHER THAN	Not relevant				
		PART OF	Not relevant				

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	FLOW (Hot gas)	***CHANGE***	Poor air connections when removing plugs in siphon or any connections from CFB to HX. (See Less).	Hot air leak or hot sand leak	(E1O) Operating procedure. (E2O) Operator detects problem during start-up		Formal sign off on operating procedure to be considered for each gasifier start. (See elsewhere) Consider automating bed pressure controls. (See elsewhere)
	FLOW (LPG)	***NONE***		No safety issues			

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		MORE	Operator error.	Unburned fuel into extraction duct	(E1SIF) CFB Oxygen controlled and cuts out LPG on low oxygen (E2MSF) High flow in extraction duct. Dilution of combustibles. See other extraction duct safeguards.		Check dilution of combustibles (See elsewhere) and **consider maximum LPG flow (JWB and RD).**
		LESS		No safety issues			
		PARTLY		Not relevant			

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	FLOW (LPG)	***AS WELL AS***	Compressed air failing. Loss of primary compressed air will cause CFB pilot failure or loss of secondary compressed air can cause slagging in CFB and pilot tube damage.	Loss of pilot flame. Buildup of unburned char.	(E1SIF) Flame supervision system. (E2O) Temperature will indicate fault to operator		

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		REVERSE	Not relevant				
		OTHER THAN	See afterburner				
		PART OF	Not relevant				
		CHANGE	See Afterburner				
	Temp.	***NONE***	Not relevant				
		MORE	See above flows				
	Temp.	***LESS***		Minor safety issues. See comment.			This will become an issue if flame supervision is altered to a photocell. (If pilot flame is not operating.)
		PARTLY	Not relevant				

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		AS WELL AS	Not relevant				
		REVERSE	Not relevant				
		OTHER THAN	Not relevant				
		PART OF	Not relevant				
		CHANGE	See other Change				

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	Power and/or Instrument Air Failure			Minor safety issues on power failure. Loss of instrument air will lead to loss of pilot. This could lead to build-up of unburned LPG	(E1SIF) Controller locks out to prevent automatic restart. **(P2O) Operator procedure for power failure and instrument air failure** (See else where)		Check behavior of services on power return. (See elsewhere) Instrument air system under review (See elsewhere). Potential interlock (See elsewhere). Consider emergency lighting (See elsewhere).

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Unit:	CAPE						
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	Flow (Air)	***NONE***	Refer to A/B Flow (air)				Start-up conditions are very similar to issues covered in CFB Node for Flow of air
		MORE	Refer to A/B Flow (air)				
		LESS	Refer to A/B Flow (air)				
		PARTLY	Refer to A/B Flow (air)				
		AS WELL AS	Refer to A/B Flow (air)				

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		REVERSE	Refer to A/B Flow (air)				
		OTHER THAN	Refer to A/B Flow (air)				
		PART OF	Refer to A/B Flow (air)				
		CHANGE	Refer to A/B Flow (air)				
	Flow (LPG)	***NONE***	Refer to CFB Flow (LPG)				

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	Flow (LPG)	***MORE***	Small reducer failure.	Unburned fuel into A/B, where it will be combusted. No safety issue	(E1SIF) BFB Oxygen controlled and cuts out LPG on low oxygen (E2MSF) A/B designed to burn combustible gas from BFB. (E3MSF) High flow in extraction duct. Dilution of combustibles. See other extraction duct safeguards.		Check dilution of combustibles and maximum LPG flow (See elsewhere).
		LESS	Refer to CFB Flow (LPG)				

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		PARTLY	Refer to CFB Flow (LPG)				
		AS WELL AS	Refer to CFB Flow (LPG)				
		REVERSE	Refer to CFB Flow (LPG)				
		OTHER THAN	Refer to CFB Flow (LPG)				
		PART OF	Refer to CFB Flow (LPG)				
		CHANGE	Refer to CFB Flow (LPG)				

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	Flow (Steam/ Hot gas)	***NONE***	Valve Failure, distributor blocked, loss of water supply	Lose steam and fluidization in BFB, chute and siphon. Possibility of air into BFB through chute. Could lead to combustible gas going to A/B and combustion occurring in BFB. These are not safety hazards.	(E1O) Pressure drop in BFB will indicate loss of fluidization. Temperatures will also show fault. (E2MSF) The BFB is designed for this occurrence. (E3MSF) The A/B is designed for this occurrence		

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		MORE	Valves fail open. Change in delivery pressure of water. Low pressure water delivery system changed over to high pressure system.	More steam fed to BFB. Lowers temperature in BFB. Possibility of carry over of particles into A/B cyclone.	(E1O) Operator action on low temperature or high flow or high pressure across bed (E2MSF) Pressure relief should pressure drop across BFB rise. (E3O) Procedure for changeover		**Check maximum possible steam flow in regards to carry over of bed material (JWB and RD)**.
		LESS	See NONE				
		PARTLY	See NONE				

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	Flow (Steam/ Hot gas)	***AS WELL AS (Air)***	Part of procedure during transition. See air in this Node				

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		REVERSE	Pressure relief vessel is empty	See A/B node but no additional BFB consequences.	(E1MSF) Hood ducting can cope with expected hot gas (&/or LPG) release. (E2O) Lack of pressure drop across BFB and low oxygen in BFB and dropping temperature in BFB. (E3SIF) LPG cuts out on low oxygen Note initiating event assumes failure of relief valve protection layer (E4SIF) Existing CO and H2 monitoring which cuts all fuels (wood and LPG) (E5MSF) Convection vents (E6MSF) Room extraction		Investigate possibility of automatic refill of pressure relief (See elsewhere).

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	Flow (Steam/ Hot gas)	***OTHER THAN***	Extremely unlikely				
		PART OF	See NONE				

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		CHANGE	Poor preliminary check of line connections. Distributor leaks. Can cause LPG leak or toxic combustible gas leak if leak isn't found during start-up	Water leak or hot (toxic and combustible) gas leak into room. See above as well.	(E1O) Rigorous cold check procedure. (E2O) Numerous instruments show leaks during start-up on air. (E3MSF) Plant ducting designed to cope with large flows (E3SIF) H2 and CO sensors cut out all fuels (E4MSF) Room extraction fans (E5MSF) Convective vents		Consider additional safeguards (LPG) (See elsewhere)

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Unit:	CAPE						
ITEM Number	**PARA METER**	**GUIDE WORD**	**CAUSES**	**CONSEQUENCES**	**PROPOSED (P) & EXISTING (E) SAFEGUARDS MSF/SIF/Other**	**SENSOR/FINAL ELEMENT Tag**	**COMMENTS ACTIONS**
	Flow (Wood)	***NONE***	Screw empty or screw has seized	Toxic and combustible gas into feed hopper. Potential fire in feed hopper.	(E1MSF) Inert gas flow into hopper and up screw (E2O) Temperature sensor in hopper and screw entry into BFB (E3MSF) Reasonably sealed hopper reduces oxygen in hopper. (E4MSF) Hopper can hold 20kg of wood pellets. (E5MSF) Vibrator on hopper to prevent loss of flow in screw while feed is in hopper.		

Design Intent:	To generate medium calorific value fuel gas from biomass in a staged reaction of steam moderated pyrolysis in a bubbling fluidized bed (BFB) and combustion of residual char in a circulating fluidized bed (CFB). The system uses flow of an inert solid (sand) to transfer heat between the two reaction stages. The fuel gas is then combusted in a downstream afterburner.						
NODE Number	BFB and Feed Delivery	**DRAWING NUMBER:**	P & I Diagram CAPE Gasifier	**NODE DESCRIPTION:**	The bubbling fluidized bed (BFB) reactor combines steam and solid biomass fuel fed from the storage hopper to form char and producer gas with heat provided by recirculating sand from the CFB.		
Unit:	CAPE						
ITEM Number	**PARA METER**	**GUIDE WORD**	**CAUSES**	**CONSEQUENCES**	**PROPOSED (P) & EXISTING (E) SAFEGUARDS MSF/SIF/Other**	**SENSOR/FINAL ELEMENT Tag**	**COMMENTS ACTIONS**
		MORE	Operator error or VF drive failure	High gas production, BFB temperature will drop.	(E1MSF) Afterburner is over designed. (E2SIF) Temperature control on A/B will cut all fuels (E3SIF) Low temperature cut out on BFB (E4O) Whole heap of stuff will indicate fault to operator.		
	Flow (Wood)	***LESS***	See none				
		PARTLY	Not relevant				

Design Intent:	To generate medium calorific value fuel gas from biomass in a staged reaction of steam moderated pyrolysis in a bubbling fluidized bed (BFB) and combustion of residual char in a circulating fluidized bed (CFB). The system uses flow of an inert solid (sand) to transfer heat between the two reaction stages. The fuel gas is then combusted in a downstream afterburner.						
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Unit:	CAPE						
ITEM Number	**PARA METER**	**GUIDE WORD**	**CAUSES**	**CONSEQUENCES**	**PROPOSED (P) & EXISTING (E) SAFEGUARDS MSF/SIF/Other**	**SENSOR/FINAL ELEMENT Tag**	**COMMENTS ACTIONS**
		AS WELL AS 1	Treated wood, wood with chlorine.	Toxic gas out of vent	(E1MSF) Vent is well designed		
		As Well As 2	Loss of purge flow	Increased risk of backflow of BFB gases. Potential burn-back and fire.	(E1O) Start-up nitrogen check and operational checks during run (E2O) Temperature sensors in hopper and screw (E3MSF) Hopper is sealed and is of small volume		
		REVERSE	Screw motor in reverse	See NONE			

Design Intent:	To generate medium calorific value fuel gas from biomass in a staged reaction of steam moderated pyrolysis in a bubbling fluidized bed (BFB) and combustion of residual char in a circulating fluidized bed (CFB). The system uses flow of an inert solid (sand) to transfer heat between the two reaction stages. The fuel gas is then combusted in a downstream afterburner.						
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Unit:	CAPE						
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		OTHER THAN	Inert gas into room from the feed hopper from operator error	Inert gas into room from the feed hopper	(E1MSF) Extraction system is over designed.		**Check automated cut-off on inert gas (JWB)**. Particularly for when gasifier isn't in operation.
		PART OF	See None				
		CHANGE	Changing feed material.	See MORE			
	Flow (sand)	***NONE***	Blockage in chute or flow failure to chute.	Not a safety issue			

Design Intent:	To generate medium calorific value fuel gas from biomass in a staged reaction of steam moderated pyrolysis in a bubbling fluidized bed (BFB) and combustion of residual char in a circulating fluidized bed (CFB). The system uses flow of an inert solid (sand) to transfer heat between the two reaction stages. The fuel gas is then combusted in a downstream afterburner.						
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Unit:	CAPE						
ITEM Number	**PARA METER**	**GUIDE WORD**	**CAUSES**	**CONSEQUENCES**	**PROPOSED (P) & EXISTING (E) SAFEGUARDS MSF/SIF/Other**	**SENSOR/FINAL ELEMENT Tag**	**COMMENTS ACTIONS**
		MORE		Not a safety issue. See blocked siphon in CFB node			
		LESS	See None				
		PARTLY	See None				
		AS WELL AS	Not relevant				
		REVERSE	Not relevant				
		OTHER THAN	See Other Flows				
		PART OF	Not relevant				
		CHANGE	See other Flow CHANGEs				

Design Intent:	To generate medium calorific value fuel gas from biomass in a staged reaction of steam moderated pyrolysis in a bubbling fluidized bed (BFB) and combustion of residual char in a circulating fluidized bed (CFB). The system uses flow of an inert solid (sand) to transfer heat between the two reaction stages. The fuel gas is then combusted in a downstream afterburner.						
NODE Number	BFB and Feed Delivery	**DRAWING NUMBER:**	P & I Diagram CAPE Gasifier	**NODE DESCRIPTION:**	The bubbling fluidized bed (BFB) reactor combines steam and solid biomass fuel fed from the storage hopper to form char and producer gas with heat provided by recirculating sand from the CFB.		
Unit:	CAPE						
ITEM Number	**PARA METER**	**GUIDE WORD**	**CAUSES**	**CONSEQUENCES**	**PROPOSED (P) & EXISTING (E) SAFEGUARDS MSF/SIF/Other**	**SENSOR/FINAL ELEMENT Tag**	**COMMENTS ACTIONS**
	Temp.	***NONE***	Not relevant				
		MORE		Not a safety issue.			
		LESS		Minor safety issues.	(E1SIF) Low temp cut-out		
		PARTLY	Not relevant				
		AS WELL AS	Not relevant				
		REVERSE	Not relevant				
BFB and feed delivery system	**Temp.**	***OTHER THAN***	Not relevant				
		PART OF	Not relevant				

Design Intent:	To generate medium calorific value fuel gas from biomass in a staged reaction of steam moderated pyrolysis in a bubbling fluidized bed (BFB) and combustion of residual char in a circulating fluidized bed (CFB). The system uses flow of an inert solid (sand) to transfer heat between the two reaction stages. The fuel gas is then combusted in a downstream afterburner.						
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ITEM Number	**PARA METER**	**GUIDE WORD**	**CAUSES**	**CONSEQUENCES**	**PROPOSED (P) & EXISTING (E) SAFEGUARDS MSF/SIF/Other**	**SENSOR/FINAL ELEMENT Tag**	**COMMENTS ACTIONS**
		CHANGE	See other Change				
	Power Failure and Instrument Air Failure		See other no Flows				**Double check logic on all systems for power failure and restart (RD)**.

Design Intent:	To generate medium calorific value fuel gas from biomass in a staged reaction of steam moderated pyrolysis in a bubbling fluidized bed (BFB) and combustion of residual char in a circulating fluidized bed (CFB). The system uses flow of an inert solid (sand) to transfer heat between the two reaction stages. The fuel gas is then combusted in a downstream afterburner.						
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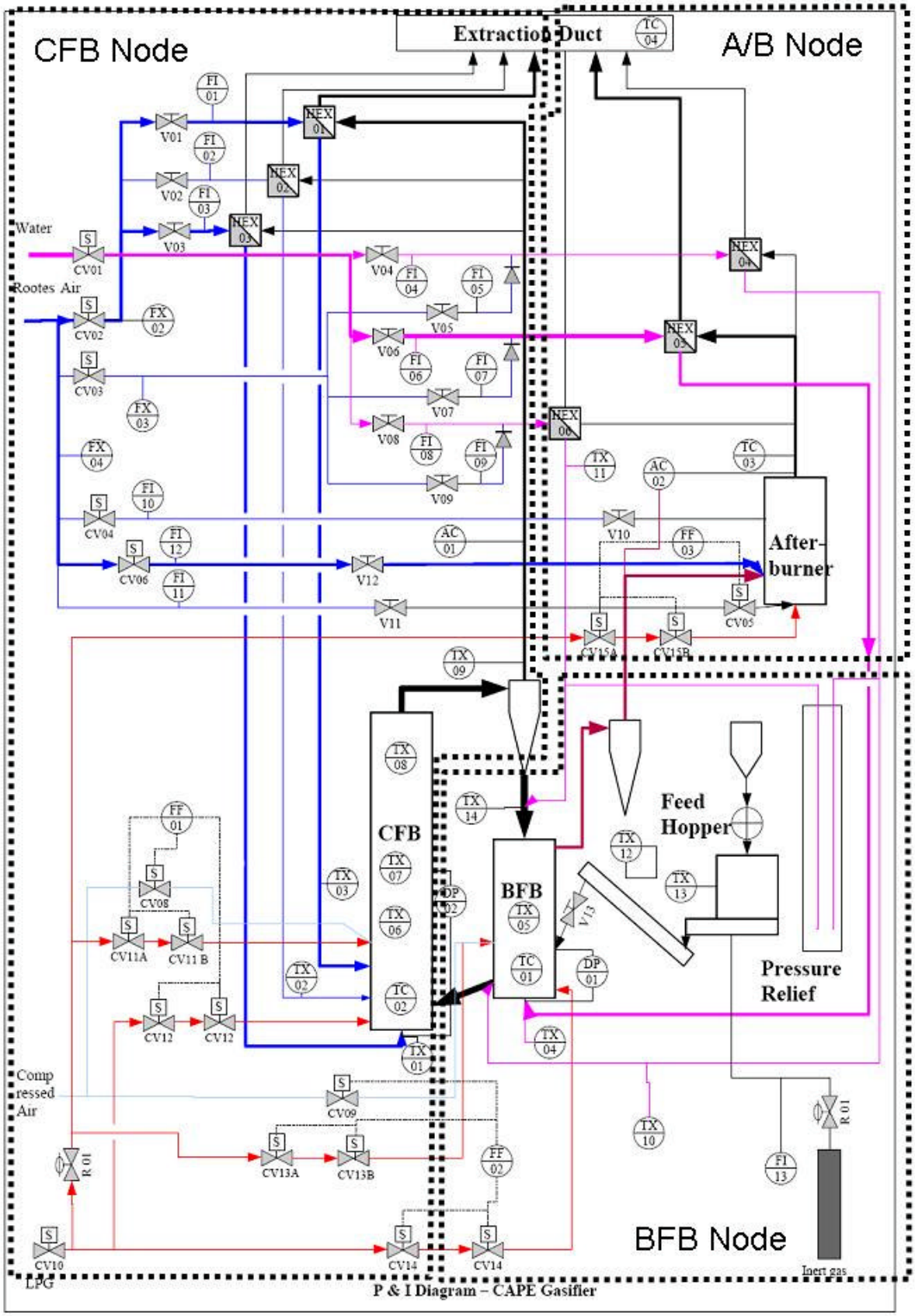
ITEM Number	**PARA METER**	**GUIDE WORD**	**CAUSES**	**CONSEQUENCES**	**PROPOSED (P) & EXISTING (E) SAFEGUARDS MSF/SIF/Other**	**SENSOR/FINAL ELEMENT Tag**	**COMMENTS ACTIONS**
Plant Extraction System Failure				A/B exhaust and CFB into room. Hot gas build-up in the duct. Possible activation of sprinkler in duct.	(E1SIF) Flow switch on extraction shuts off process. (E2MSF) Room extraction fans (E3MSF) Convection vents (E4SIF) CO Sensor and H2 sensor (E5O) Visual indication of status of fire hood in control room		**Add fire hood fan speed to operating log (JWB).** **Check activation temperature of hood sprinkler and report to procedure (DB)** **Consider feedback signal from fan (RD & BG).**

Nodes marked on P&ID for HAZOP:



F. Appendix F

HPLC Tar Analysis and Infra Red Spectra

This section includes the HPCL chromatograms for the standard tar mix, which does not contain a complete set of the compounds in the tar material. The standard mix was used to identify the main compounds in the tar. Infra-red spectroscopy was used by Sean Devenish to make a library of these compounds for future identification however this library is not yet complete. Figure 0-1 shows the HPLC chromatogram for the standard mix of tar compounds. The following figures are a complete set of the IR spectra for all the tar compounds identified.

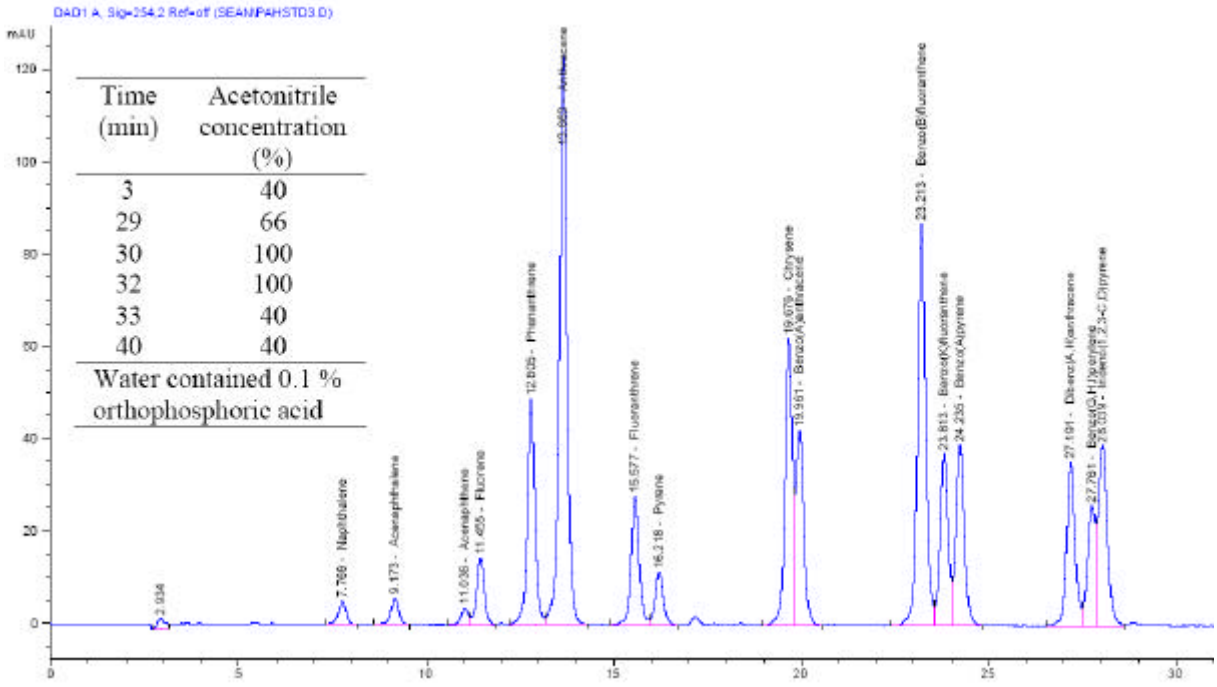
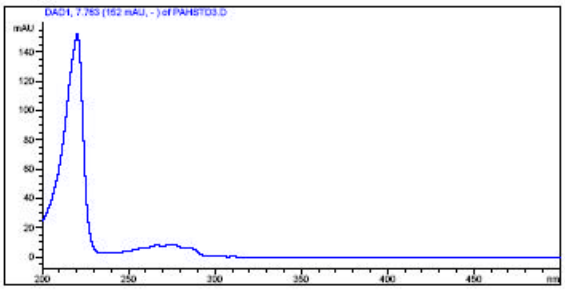
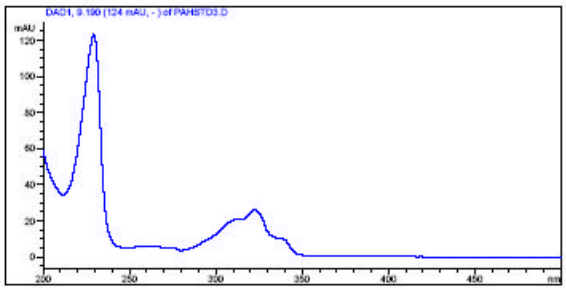


Figure 0-1. HPLC separation of standard mix of tar compunds.

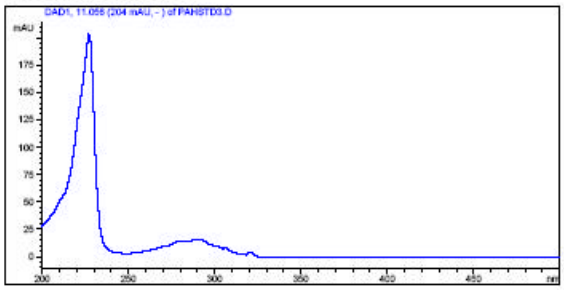
Napthalene



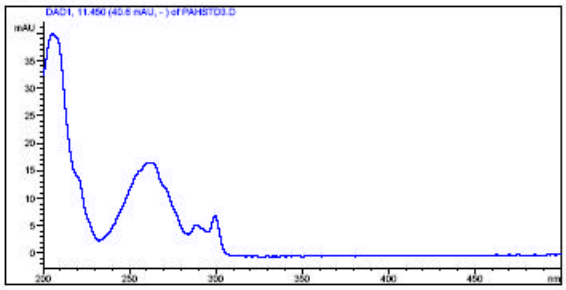
Acenaphthalene



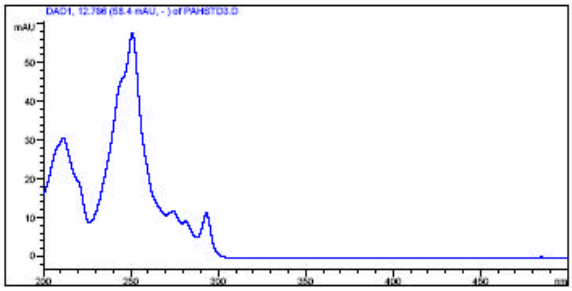
Acenaphthene



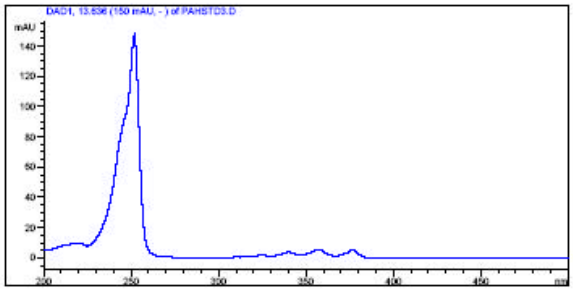
Fluorene



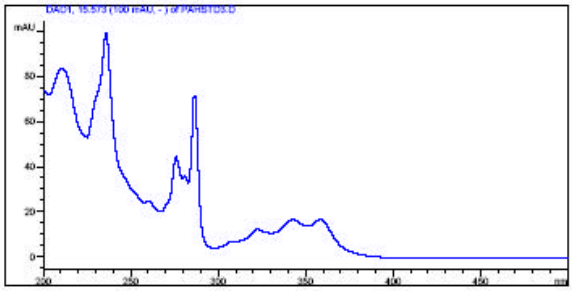
Phenanthrene



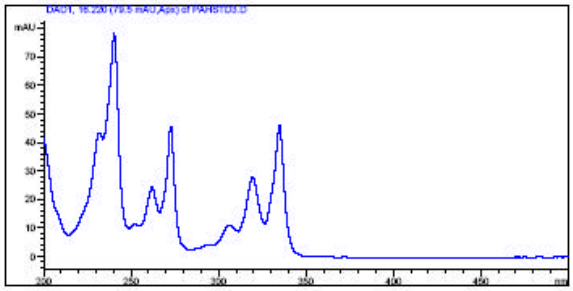
Anthracene



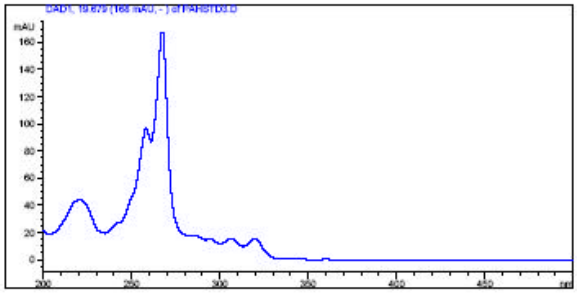
Fluoranthrene



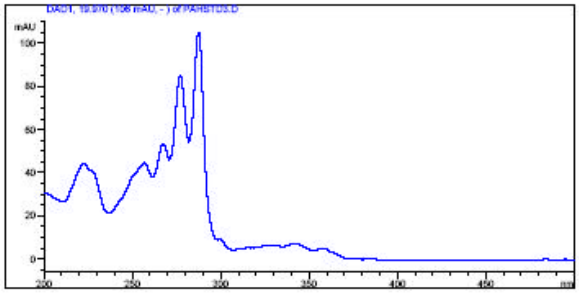
Pyrene



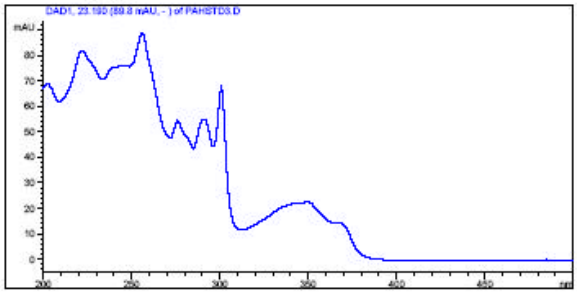
Chrysene



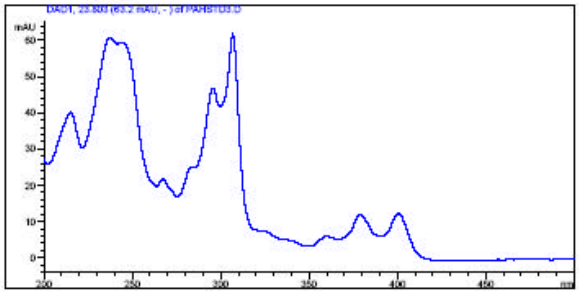
Benzo(A)anthracene



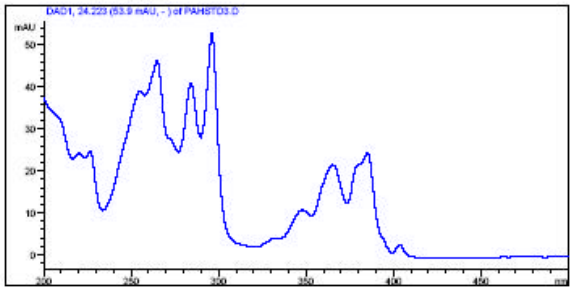
Benzo(B)fluoranthene



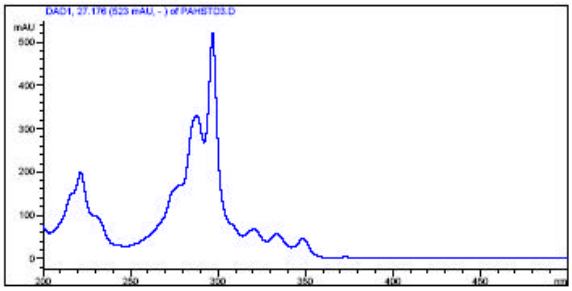
Benzo(K)fluoranthene



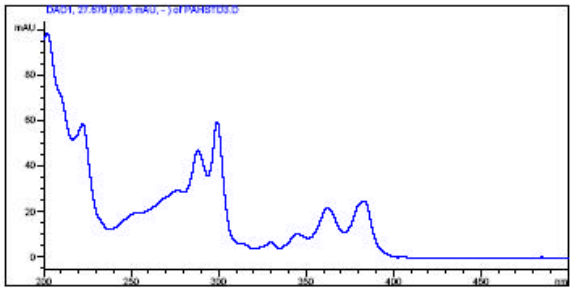
Benzo(A)pyrene



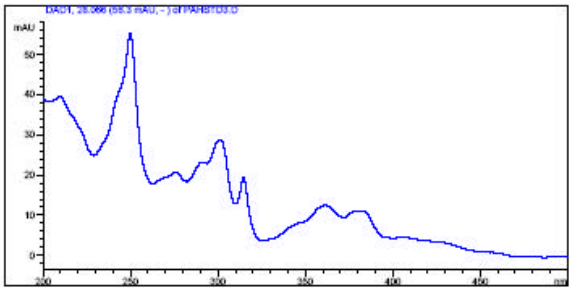
Dibenz(A,H)anthracene



Benzo(G,H,I)perylene



Indeno(1,2,3-C,D)pyrene



G. Appendix G

Micro GC Chromatograms

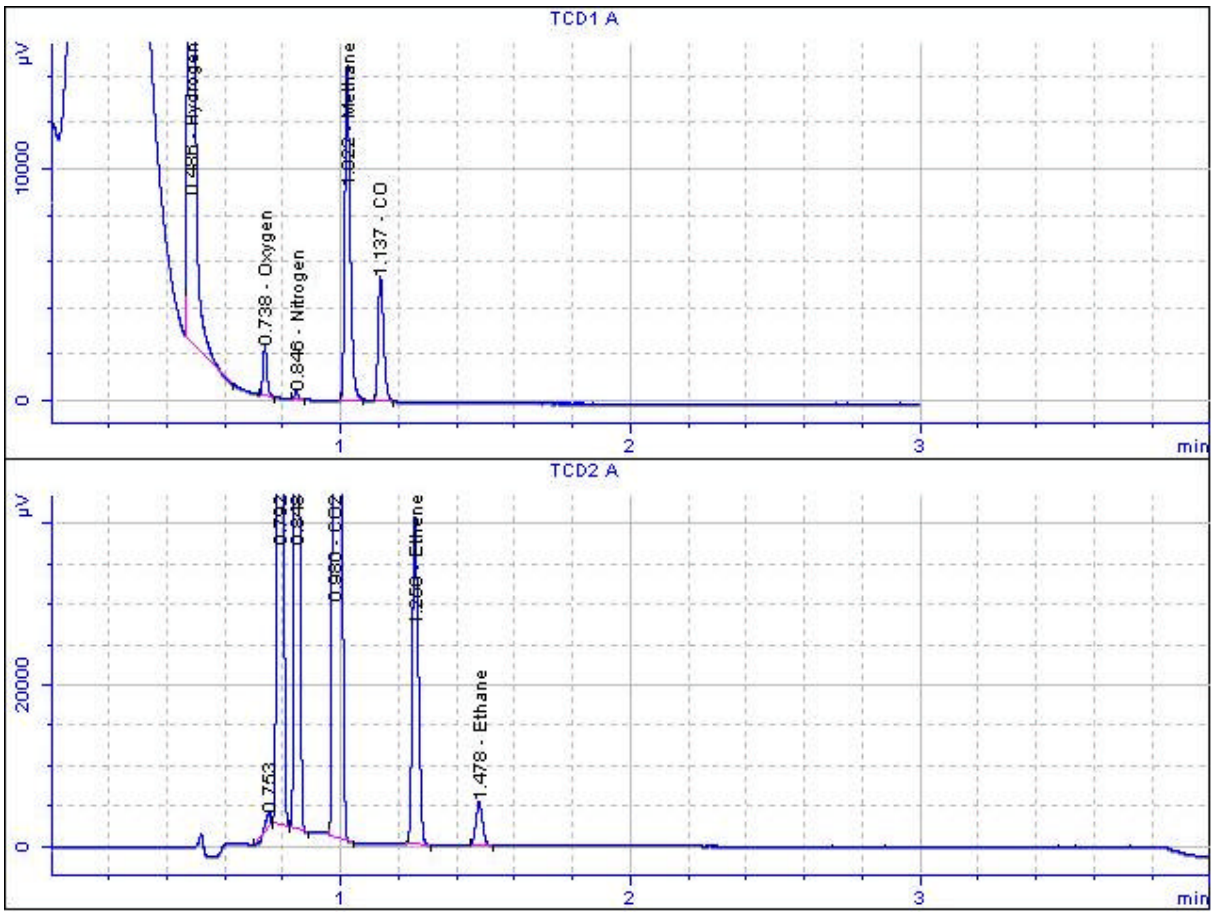


Figure 0-1. Chromatogram of calibration gas

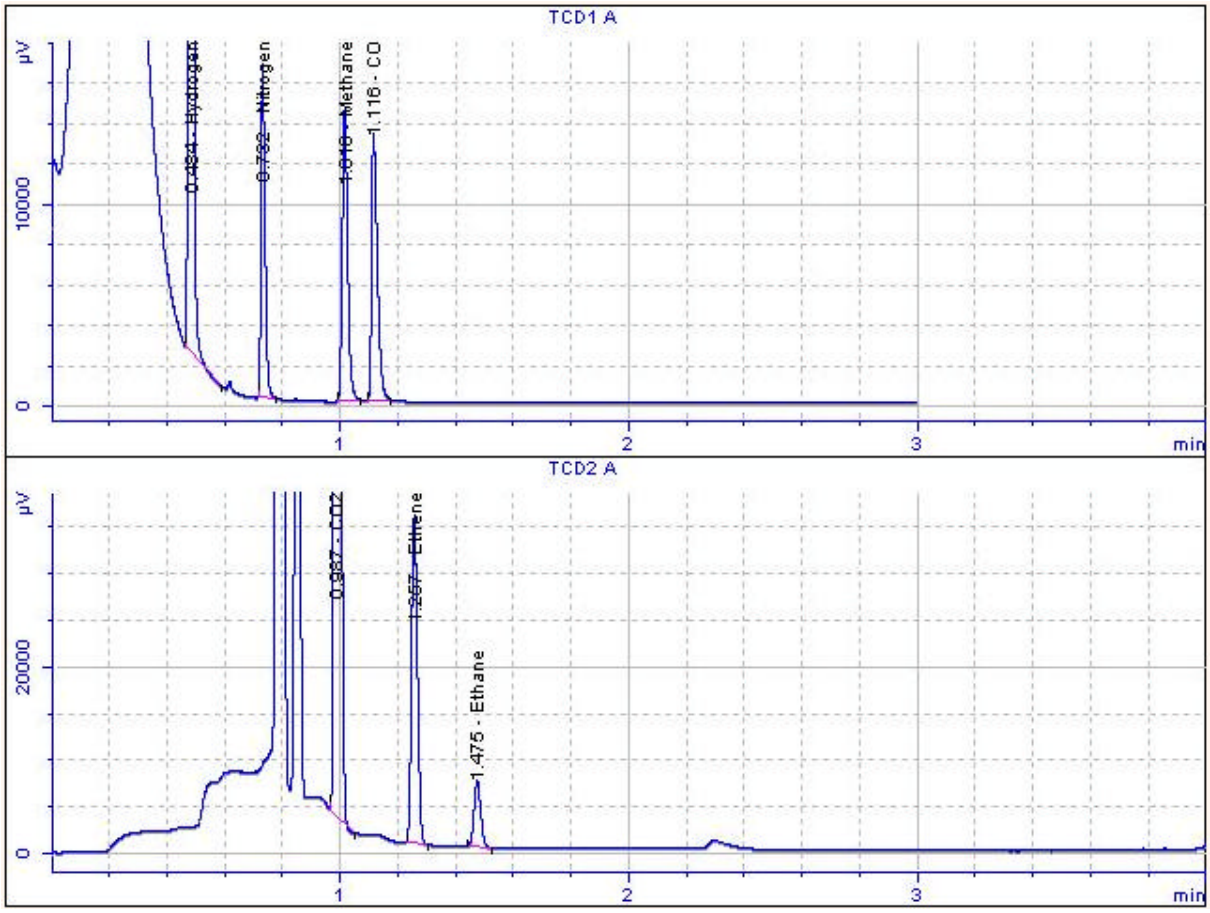


Figure 0-2. Chromatogram of CAPE gasifier product gas

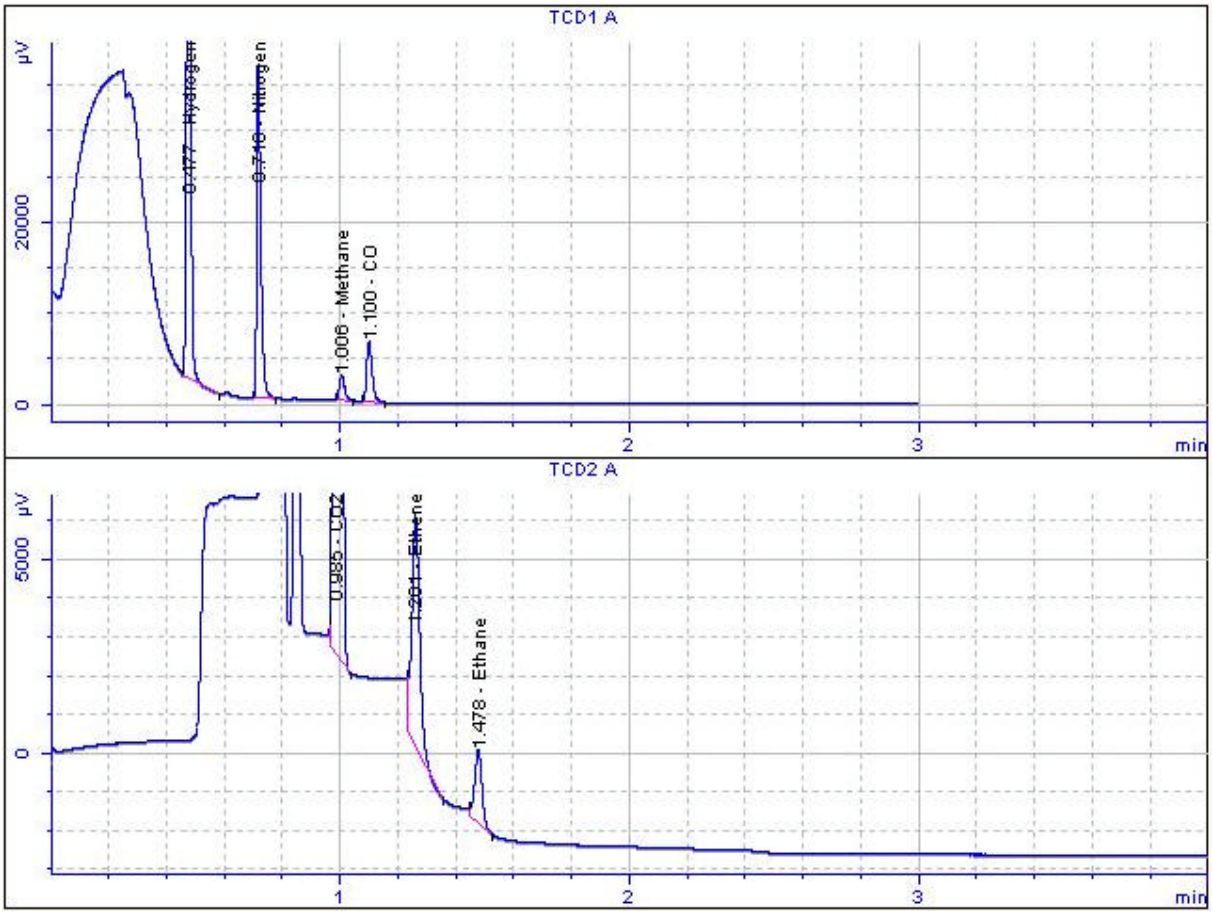


Figure 0-3. Chromatogram of Page Macrae gasifier product gas.